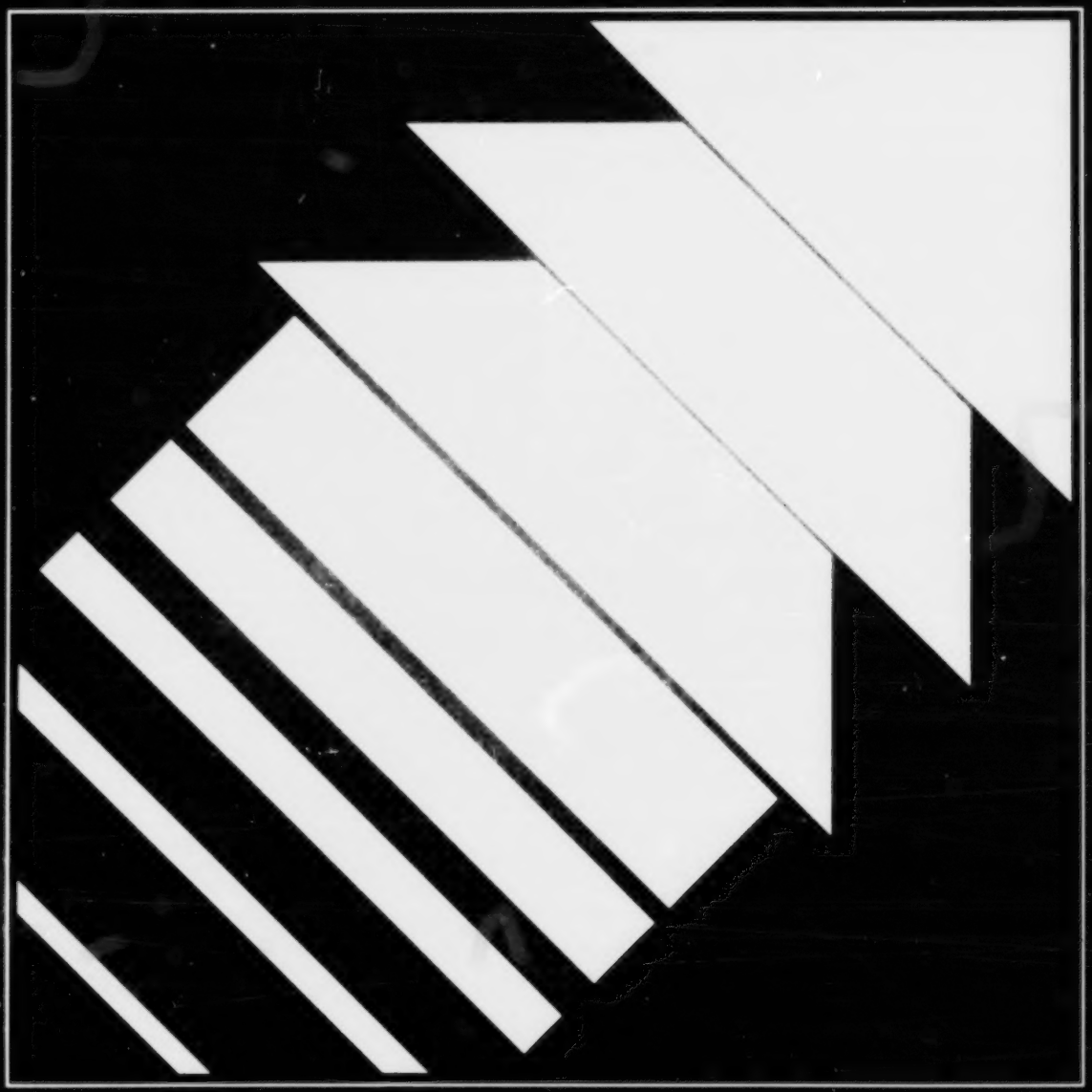
NATIONAL URANIUM TAILINGS PROGRAM



Energy, Mines and Resources Canada
Energie, Mines et Ressources Canada

Canada

ALTERNATIVE URANIUM MILL PROCESSING TECHNOLOGIES

5-1732

DSS File #15SQ.23241-~~4-1694~~
Contract #OSQ84-00414

A Research Report Prepared for:

The National Uranium Tailings Program
Canada Centre for Mineral and Energy Technology
Energy, Mines and Resources Canada
555 Booth Street
Ottawa, Ontario
K1A 0G1

by

SENES Consultants Limited
499 McNicoll Avenue
Willowdale, Ontario
M2H 2C9

March 1986

129 pp

JUL 10 1986
MICROMEDIA

TABLE OF CONTENTS

		Page
1.0	EXECUTIVE OVERVIEW AND SUMMARY	
	1.1 Background	1-1
	1.2 Study Results	1-2
2.0	INTRODUCTION	
	2.1 Background	2-1
	2.2 Study Objectives/Terms of Reference	2-2
	2.3 Study Basis	2-2
	2.4 Report Contributions	2-3
3.0	PROCESS DESCRIPTIONS AND ALTERNATIVE SELECTIONS	
	3.1 Overview	3-1
	3.2 Sulphuric Acid Leaching and Process Modification	3-2
	3.2.1 Sulphuric Acid Leach - Base Case	3-2
	3.2.2 Pyrite Flotation, Acid Manufacture	3-3
	3.2.3 Radioisotope Flotation	3-5
	3.2.4 Thorium Recovery	3-5
	3.2.5 Pressure Leach	3-7
	3.2.6 Radium Extraction from Tailings	3-8
	3.3 Hydrochloric Acid Leach	3-10
	3.4 Chlorination Process	3-12
	3.5 Carbonate Leach	3-13
	3.6 Other Milling Alternatives	3-15
	3.6.1 Heap Leaching	3-15
	3.6.2 Nitric Acid Leach	3-16
	3.6.3 Preconcentration	3-18
	3.6.4 In-Situ Leaching	3-19
	3.6.5 Ferric Chloride and Ferric Sulphate Leaching	3-20
	3.6.6 Caro's Acid and Presulphuric Acid Leaching	3-20
	3.7 Selected Process Alternatives	3-20

4.0 DEVELOPMENT OF PROCESS FLOWSHEETS AND COST ESTIMATES

4.1	General	4-1
4.1.1	Design Capacity	4-1
4.1.2	Costing Basis	4-1
4.1.3	Capital Costs	4-2
4.1.4	Operating Costs	4-2
4.1.5	Process Flowsheets	4-3
4.2	Process Description and Cost Estimates for Ore A Milling Alternatives	4-3
4.2.1	Base Case - Sulphuric Acid Leach	4-3
4.2.2	Alternative 1 - Pyrite Flotation	4-7
4.2.3	Alternative 2 - Acid Production	4-10
4.2.4	Alternative 3 - Radioisotope Bulk Flotation	4-13
4.2.5	Alternative 4 - Thorium Recovery	4-15
4.2.6	Alternative 5 - Autogenous Leach	4-18
4.2.7	Alternative 6 - EDTA Leach	4-21
4.2.8	Alternative 7 - HCl Leach	4-24
4.3	Process Descriptions and Cost Estimates for Ore B Milling Alternatives	4-28
4.3.1	Base Case - Sulphuric Acid Leach	4-28
4.3.2	Alternative 1- HCl Leach	4-32
4.3.3	Alternative 2 - HCl + CL_2 Leach	4-35
4.3.4	Alternative 3 - Carbonate Leach	4-39
4.4	Process Descriptions and Cost Estimates for Ore C Milling Alternatives	4-42
4.4.1	Base Case - Sulphuric Acid Leach	4-42
4.4.2	Alternative 1 - HCl Leach	4-46
4.4.3	Alternative 2 - HCl + Cl_2 Leach	4-48
4.5	Discussion and Summary	4-52
4.5.1	General	4-52

5.0 HEALTH AND SAFETY

5.1 Introduction 5-1
5.2 Ore A - Process Alternatives 5-1
5.2.1 Alternative 1 - Pyrite Flotation 5-1
5.2.2 Alternative 2 - Acid Production 5-2
5.2.3 Alternative 3 - Bulk Flotation 5-2
5.2.4 Alternative 4 - Thorium Recovery 5-3
5.2.5 Alternative 5 - Autogenous Leach 5-3
5.2.6 Alternative 6 - EDTA Leach 5-3
5.2.7 Alternative 7 - HCl Leach 5-4
5.3 Ore B - Milling Alternatives 5-5
5.3.1 Alternative 1 - HCl Leach 5-5
5.3.2 Alternative 2 - HCl + Cl_2 Leach 5-5
5.3.3 Alternative 3 - Carbonate Leaching 5-6
5.4 Ore C - Milling Alternatives 5-6
5.4.1 Alternative 1 - HCl Leach 5-6
5.4.2 Alternative 2 - HCl + Cl_2 Leach 5-6

6.0 ENVIRONMENTAL ASPECTS

6.1 Introduction 6-1
6.1.1 Overview 6-1
6.1.2 Tailings and Effluent Management Requirements - Base Case 6-1
6.1.3 Close-out Requirements - Base Case 6-3
6.1.4 Environmental and Radiological Concerns 6-5
6.2 Characteristics of Solid and Liquid Discharges 6-6
6.2.1 Solid Discharges 6-6
6.2.2 Liquid Discharges 6-9
6.3 Estimated Costs 6-10
6.3.1 Tailings Disposal and Effluent Treatment Operating Costs 6-10
6.3.2 Abandonment Costs 6-12
6.3.3 Long-Term Treatment Costs 6-12
6.3.4 Total Costs for Tailings Management 6-13

6.4 Comparison of Alternatives 6-14
6.4.1 General 6-14
6.4.2 Type A Ore Milling Alternatives 6-14
6.4.3 Type B Ore Milling Alternatives 6-16
6.4.4 Type C Ore Milling Alternatives 6-17

7.0 SUMMARY
7.1 Overall Costs 7-1
7.2 Type A Ore Alternatives 7-2
7.3 Type B Milling Alternatives 7-4
7.4 Type C Milling Alternatives 7-4

8.0 REFERENCES

LIST OF FIGURES

Figure		Follows Page
4.2.1	Base Case Sulphuric Acid Leach	4-5
4.2.2	Alternative 1 - Pyrite Flotation	4-8
4.2.3	Alternative 2 - Acid Production	4-11
4.2.4	Alternative 3 - Bulk Flotation	4-13
4.2.5	Alternative 4 - Thorium Precipitation	4-16
4.2.6	Alternative 5 - Autogenous Leach	4-19
4.2.7	Alternative 6 - EDTA Leach	4-22
4.2.8	Alternative 7 - HCl Leach	4-26
4.3.1	Base Case - Sulphuric Acid Leach	4-29
4.3.2	Alternative 1 - HCl Leach	4-33
4.3.3	Alternative 2 - HCl + Cl_2 Leach	4-36
4.3.4	Alternative 3 - Carbonate Leach	4-40
4.4.1	Base Case - Sulphuric Acid Leach	4-43
4.4.2	Alternative 1 - HCl Leach	4-46
4.3.3	Alternative 2 - HCl + Cl_2 Leach	4-50

LIST OF TABLES

Table		Follows Page
1.1	Overall Cost Summary	1-3
4.5.1	Capital and Operating Cost Summary for Ore Milling Alternatives	4-52
6.2.1	Estimated Chemical and Radiological Characteristics - Type A, B & C Ore and Tailings	6-6
6.2.2	Summary of Concentrate Volumes & Characteristics	6-6
6.2.3	Typical Chemical Characteristics Tailings Effluent Discharge	6-9
6.3.1	Operating Costs for Tailings Disposal and Effluent Treatment	6-10
6.3.2	Abandonment Cost (w/o Treatment)	6-12
6.3.3	Long-Term Treatment Costs (10-200 Years After Closure)	6-13
6.3.4	Total Costs Tailings Disposal and Effluent Management	6-13
6.4.1	Comparison of Alternative Processes to the Base Case From an Environmental Perspective	6-14
7.1.1	Overall Cost Summary	7-1
7.1.2	Factors in Overall Cost	7-1

1.0 EXECUTIVE OVERVIEW AND SUMMARY

1.1 Background

During the mid 1970's to early 1980's there was a flurry of research into environmentally-superior uranium mill processing alternatives. During this period, uranium world prices soared to more than \$100/kg U_3O_8 and uranium exploration, milling expansions and new mine developments were highly active. Given the long term prediction for high uranium demand and considering the high world prices, there was a general concensus that environmentally more acceptable processes would be economically justified. With the drop in uranium prices to the \$50/kg U_3O_8 range by the mid 1980's the general view as to the economics of alternative processes has obviously been altered.

There are three primary ore types which have been milled in Canada. These are the low-grade (0.1% U_3O_8) sulphidic Elliot Lake types ores, the medium-grade (0.2 to 0.4 % U_3O_8) low heavy metal content Rabbit Lake ores and the high-grade (>2% U_3O_8) nickel-cobalt-arsenide Key Lake type ores. In this report these have been referred to type A, B and C ore respectively. All of these ores have been processed by modifications of the conventional sulphuric acid leaching process, using solvent extraction or ion exchange for product concentration and various precipitating agents, of which ammonia and magnesium are most common.

The expansion and development of uranium projects from 1976 onwards has been conducted under an environmental microscope. During that period the Cluff Lake Board of Inquiry 1977/78, the Environmental Assessment of the Elliot Lake Mining Expansion 1977-1979 and the Key Lake Board of Inquiry 1980/81 and other studies took place which for some lead to mining development while for others resulted in several commissions and assessment boards ultimately leading to moratoriums and cancelled projects.

During these hearing and assessments, process alternatives were reviewed in detail. Based upon these reviews, sulphuric acid leaching was retained as the

principal processing option, however several modifications were reviewed. The only primary change to milling processes occurred in Elliot Lake where the rehabilitated facilities at Panel and Stanleigh adopted alternative stripping and precipitation agents for environmental reasons. These were both operational changes which would have little effect on the long-term characteristics of tailings.

In Saskatchewan the trend was to develop tailings repositories to better manage the tailings than to modify mills to produce "cleaner tailings". This resulted in the concepts of concrete vault disposal for high grade residues at Cluff Lake, engineered above ground disposal and encapsulation at Key Lake and recently, in pit disposal for the Collins 'B' development.

The Elliot Lake operations seriously considered pyrite removal and acid production but for these low grade ores; however, the economics have repeatedly demonstrated this alternative cannot be justified.

More recently operators have become increasingly concerned with the costs for the decommissioning of their properties and more specifically the uranium tailings. Given that mining operations have experienced major expenditures for closing-out their properties, the question now arises - are there environmentally superior process alternatives that can be justified when consideration is given to the reduced long-term tailings management costs and environmental benefits? This study has tried to identify potential environmentally superior processes and has assessed their overall costs and benefits.

1.2 Study Results

The following are the major findings of this study:

- without exception the alternative milling processes have higher overall costs

- processes which produce "cleaner tailings" produce concentrates which are expensive to dispose of and may present an equivalent or greater hazard.
- perpetual treatment would appear to be the simplest and most cost efficient long-term management plan for type A and B ores
- process alternatives for type A ore such as pyrite removal or bulk flotation have many benefits. These processes have low capital costs but high operating costs. Therefore, although they may be uneconomic for implementation at the startup of an operation they may prove to be more applicable during the latter years of operation. This could serve to place a cover of pyrite reduced or radionuclide-reduced tailings over the main tailings deposit.
- estimated total waste management costs for tailings disposal and long-term management are significant for all operations ranging from 5 per cent of the processing cost for type B ($0.90/kg U_3O_8) ore to 12.5 percent of the processing cost for type C ore.
- estimated long-term waste management costs are significant. The base case cost is $1.30/kg U_3O_8 for type A ore, $0.50/kg U_3O_8 for type B ore and $1.05/kg U_3O_8 for type C ore. In total dollars, over the twenty year lifetime of the facility assumed in this report, this is approximately $60 million for type A ore, $23 million for type B ore and $48 million for type C ore.
- only one process for type A ore and no process for type B and C ore was identified that produced "clean tailings". There will likely be institutional controls required for even the process which produce the supposedly "cleaner tailings" materials.
- the overall cost summary is provided in Table 1.1.

Table 1

OVERALL COST SUMMARY

Ore Type	Alternative	Production (10^6 kg U_3O_8/a)	Capital Cost ($/kg U_3O_8)	Operating Cost ($/kg U_3O_8)	Disposal Cost ($/kg U_3O_8)	Total Cost ($/kg U_3O_8)	Per cent Difference
A	BASE CASE	2.26	7.93	15.29	2.22	25.44	-
A1-	Pyrite Flotation	2.26	8.46	16.72	2.29	27.48	+8
A2-	Acid Production	2.26	10.82	15.93	1.93	28.68	+13
A3-	Bulk Flotation	2.26	8.75	19.12	4.31	32.18	+26
A4-	Thorium Precipitation	2.26	8.79	16.97	2.92	28.68	+13
A5-	Autogenous Leach	2.26	10.38	16.10	2.15	28.63	+13
A6-	EDTA Leach	2.26	12.31	40.26	2.74	55.31	+117
A7-	HCl Leach	2.33	14.87	30.48	3.11	48.46	+90
B	BASE CASE	2.26	6.39	10.58	0.90	17.80	-
B1-	HCl Leach	2.31	9.27	16.62	1.02	26.91	+51
B2-	HCl + Cl_2 Leach	2.26	13.23	22.41	0.73	36.37	+104
B3-	Carbonate Leach	2.26	6.50	11.01	0.47	17.98	+1
C	BASE CASE	2.26	5.73	6.66	1.77	14.16	-
C1-	HCl Leach	2.27	6.54	12.75	0.95	20.24	+43
C2-	HCl + Cl_2 Leach	2.20	7.87	15.70	0.50	24.07	+70

Ore C - A high grade arsenical ore typical of the Athabascan sandstone ore found in several northern Saskatchewan ore bodies.

The hypothetical projects at which these ores would be mined and milled have been treated as new developments, with the following basis production criteria chosen by A.H. Ross & Associates:

Ore Type	Grade % U_3O_8	Production 10^6 kgU_3O_8/a
A	0.09	2.26
B	0.40	2.26
C	2.00	2.26

Other ore constituents selected for the study are:

	Ore A	Ore B	Ore C
Thorium, %	0.025	0.002	0.02
Radium, Bq/g*	9.26	41.5	207.0
Arsenic, %	---	0.02	1.5
Nickel, %	---	---	2.0
Sulphide sulphur, %	3.0	0.1	1.0

* ores are assumed to be in secular equilibrium with U-238

2.4 Report Contributions

The contract for this study was awarded to A.H. Ross & Associates who formally ceased business operations in November 1985. Prior to that time, A.H. Ross & Associates had submitted a partial draft of the report for review. This draft included a review of the process alternatives, detailed flowsheets and costing for each alternative, a review of the health and safety aspects of the process alternatives, and a section on environmental considerations.

In January 1986 SENES Consultants Limited was awarded the contract to revise and update the draft prepared by A.H. Ross & Associates and to prepare a final report. In light of these facts Chapters 3 and 4 are edited versions of the original material prepared by A.H. Ross & Associates. Chapter 5 is a revised and expanded version of the original material prepared by Mr. Peter Pullen, a private consultant. Chapter 6 is a revised version of the material originally prepared by SENES based upon data supplied by A.H. Ross & Associates. Chapters 1 and 7 are new summaries prepared by SENES.

3.0 PROCESS DESCRIPTIONS AND ALTERNATIVE SELECTIONS

3.1 Overview

Environmental considerations have played a major role in the uranium mining industry. These considerations have dealt with both the radiological and non-radiological components of the ore, the reagents employed in milling, the actual milling process selected, the design and operation of the tailings management system, and the long-term abandonment concepts.

Prior to the mid 1970's the selection of uranium milling alternatives was based primarily upon economics. Reagents like nitric acid and ammonia were utilized because they were efficient and cost effective. When concerns were raised over nitrogen loadings and ammonia toxicity, alternative processes were introduced to either eliminate these compounds or reduce the loss of the constituent to the environment. Processes such as "LAMIX" were introduced at some locations and ammonia recovery circuits installed at others. A major concern was also developing related to the long-term control of acidity and the long-term abandonment of the thousands of acres of uranium tailings. As a result of these concerns, several processes were suggested and others developed to provide for better long-term management of the tailings. Some of these involved control of acidity, while others were directed at reducing the levels of residual radioactivity in the tailings to produce a "clean tailings."

Two basic milling processes have been utilized in Canada. These are sulphuric acid leaching and carbonate leaching. In addition to these basic leaching alternatives, several other alternatives have been suggested and in some cases implemented elsewhere. These include nitric acid leaching, hydrochloric acid leaching and chlorination. These processes have been suggested as alternatives which could reduce the radium content of the tailings. Other than direct milling of ores, heap leaching and underground leaching of ores have also been attempted with varying degrees of success.

This section discusses the major process alternatives for the milling of Canadian uranium ores with an emphasis on those processes which could produce environmentally more acceptable tailings. Based upon a review of this information, altern tives were selected for each ore type for costing and environmental review.

3.2 Sulphuric Acid Leaching and Process Modifications

3.2.1 Sulphuric Acid Leach - Base Case

Except for a limited number of ores containing excessive amounts of acid consuming gangue material, uranium is universally extracted from its ores with sulphuric acid. The flowsheet starts with crushing run-of-mine ore to a size suitable for grinding. The pulp from grinding is often thickened, and sometimes filtered before entering the leach system. Uranium is extracted from the ore with a solution of sulphuric acid, usually with the help of an oxidant. The conditions of leach time, temperature, pulp density, grind size, acidity, and emf are chosen to suit the particular uranium and gangue minerals of the feed ore and the economic circumstances.

At the discharge of the leach system the solids and liquid are separated using filter or CCD systems preparatory to the discard of the solids and uranium recovery from the solution. In order to minimize the loss of soluble uranium with the solids, wash water must be added, resulting in a dilution of the solution going to the uranium recovery circuit.

Uranium recovery from solution is achieved by strong-base resin ion exchange or, as in most North American plants built since 1958, by solvent extraction using a weak base extractant. To obtain good resin loading the solution pH should not be below about 1.8 and the addition of lime to increase the pH may be required. Contrariwise, solvent loading tends to decrease above a pH of 1.8-2.0, and acid addition to the solution may be required in some cases. Because ion exchange is generally considered more applicable to low grade solution and is universally employed in the Elliot Lake uranium plants it has

been chosen for Ore A. The solvent extraction process, however, has also been chosen for Ore A because:

- the use of ion exchange in Elliot Lake is largely a result of historical timing rather than comparative evaluation.
- current Elliot Lake (ion exchange) flowsheets neutralize most of the free acid leaving the leach circuit, making the reuse of this acid impossible. Solvent extraction enables the straightforward recycle of some 4 kg acid/ton ore and this saving (and equivalent limestone costs) is regarded as decisive.

Numerous chemical systems can be employed for stripping uranium from the solvent and its subsequent precipitation as yellowcake. In the absence of interfering metals (e.g. molybdenum), ammonia systems perform satisfactorily and usually provide the best economics; but in some jurisdictions, the accompanying ammonium sulphate cannot be discharged for environmental reasons. Its recovery as a fertilizer grade product is practised at one location (Key Lake).

The washed solids and excess plant solutions are sent to the tailings pond after neutralization with lime and/or limestone to pH 8.0, or higher depending on the composition of the ore. Radium in liquid effluent is reduced to an acceptable level by treatment with barium chloride; if necessary the pH is also adjusted. A more elaborate tailings treatment system may be required depending on the composition of the ore.

3.2.2 Pyrite Flotation, Acid Manufacture

Flotation of pyrite after leaching is an important possible modification of the conventional acid leach process for Ore A because acid generation in the tailings storage pond is greatly reduced by removal of the sulphide-bearing material from the tailings. Special provision is required for disposal of the pyrite concentrate. Alternatively, the pyrite can be converted to sulphuric acid for internal mill use and for sale of any excess. Flotation of pyrite

from leach tailings eliminates the loss of uranium which would occur with flotation of pyrite from the ore.

This process modification is not considered significant for type B and C ores.

The process constitutes an 'addition' to the conventional atmospheric sulphuric acid leach flowsheet which is unchanged. The washed leach discharge solids are the feed to flotation, and the flotation tailings are neutralized for regular disposal.

CANMET conducted laboratory and small scale continuous flotation tests on samples of Dension tailings during 1976-1977 (1) and Dension Mines operated a 1-1.5 tph pilot plant on current Denison tailings for a 3 week period in September 1978 (2). Some uranium, radium (Ra-226), and thorium floated with the pyrite.

CANMET upgraded some pyrite rougher concentrate from the continuous test runs by regrinding and reflotation to make a product more suitable for the manufacture of sulphuric acid (1). The cleaning step reduced the levels of thorium and radium in the concentrates.

For this study, 90% removal of pyrite from the tailings has been assumed. Provision has been included for upgrading the rougher concentrate to 48% S to minimize the quantity of material for special disposal and to enhance the roasting-acid manufacture process.

Possible disposal methods for pyrite concentrate include:

- separate surface (underwater) disposal;
- blending into mine backfill;
- disposal with the main tailings, but using special deposition techniques to locate the sulphides at the bottom of the pile where oxidation will be very slow.

3.2.3 Radioisotope Flotation

Flotation of radium from sulphuric acid leach tailings, and separate disposal of the concentrate has been considered.

Testwork by CANMET (1) and Denison Mines Ltd. (2,3) has demonstrated approximately 65% overall removal of radium (Ra-226) from type A ore acid leach tailings by pyrite flotation (Section 3.1.2) followed by radioisotope flotation using a fatty acid collector. Combined concentrate weight was about 30% of tailings feed to flotation. The data also indicate approximately 65% recovery of thorium to the combined concentrates.

Radium and thorium contents in the flotation tailings were on the order of 4.4 Bq/g Ra-226 and 0.015% Th. Tailings containing as little as 1.9 Bq/g Ra-226 were obtained in specific tests, but combined concentrate weight was higher at about 39%.

The following tabulation illustrates typical results from the Denison pilot plant testwork.

Analyses	S (%)	U (%)	Ra-226 (Bq/g)	Th (%)
Flotation Feed	3.2	0.006	9.6	0.030
Flotation Tailings	0.1	0.004	4.6	0.015
% Recovery in Combined Concentrates (30% weight)	98	55	65	65

3.2.4 Thorium Recovery

Although a number of isotopes of thorium exist, only two are of significance for purposes of this study. Th-230 is a product of the U-238 decay series and is the precursor to Ra-226. Th-232 has a very long half-life (1.39×10^{10}

years), and is the precursor to Ra-228. Ore A contains about 0.025% total thorium, and approximately 0.005% of the total thorium is Th-230. For this situation, the relative activity of Ra-228 is about 10% of the activity of Ra-226 in the ore. Moreover, Ra-228 has about one-half the radiological toxicity of Ra-226. Thus, Th-232 is of minor environmental importance relative to Th-230 despite the fact that it constitutes most of the total quantity of thorium in the ore.

Of the three ore types, Ore A contains the greatest amount of total thorium, but the lowest concentration of Th-230 (because the uranium grade is the lowest). Thus, if Th-230 is an environmental concern over the long-term because of its Ra-226 decay product, then the concern applies to all three ores. However, there are no data relative to isotope deportment for B and C type ores, and no attempt has been made to isolate the Th-230 from the conventional sulphuric acid leaching circuits for these ores.

With Ore A, approximately 60% of the total thorium is dissolved during conventional sulphuric acid leaching. However, because of the association of Th-230 with uranium, it is possible that the Th-230 may be preferentially extracted, and for this report Th-230 dissolution has been assumed equal to uranium dissolution.

Most of the dissolved thorium remains in the raffinate from solvent extraction and is eventually advanced to tailings neutralization where it is precipitated. It is discharged to tailings as solids which are in a chemically stable state provided the pH in the tailings pond remains greater than 4 (4).

Approximately 94% of the thorium in the uranium solvent-extraction raffinate can be recovered by a solvent extraction process using a mixture of DEMPA and TBP as solvent. The thorium is stripped by 10 N sulphuric acid and simultaneously precipitated as thorium sulphate (5).

Thorium sulphate was produced commercially at Rio Algom by this process during the period 1959-1968. The plant was shut down when the market declined.

Based on the plausible assumption of preferential Th-230 dissolution, it would appear that approximately 90% of the Th-230 can be isolated in the thorium sulphate product, and thus removed from the main tailings.

3.2.5 Pressure Leach

Sulphuric acid pressure leaching is an environmentally significant modification to the conventional atmospheric leach flowsheet for Ore A because acid generation in tailings, a major concern in Elliot Lake, is largely eliminated. Moreover, assuming a sulphide sulphur level in Ore A that is similar to existing Elliot Lake ores, the oxidation of pyrite in the autoclaves is sufficient to satisfy the acid requirements of the ore.

In the case of Ore B, pressure leaching does not materially change the environmental picture associated with the conventional flowsheet, and to our knowledge has not been tested.

Low temperature pressure leaching has been selected as the base case for Ore C for this study because it provides the oxidation conditions required for uranium extraction without the use of chemical oxidants. This process alternative is discussed in Chapter 4.

Pressure acid leach testwork has been conducted on a number of Elliot Lake ores (type A ore) by CANMET (7,5), and by an operating company in the area. Those investigations were concerned primarily with conditions promoting uranium extraction rather than with achieving maximum reaction of the sulphide sulphur. Sulphur extraction up to 60% was reported and the sulphur reaction products were characterized (8). In general, uranium extractions exceeding 95% were achieved in a matter of 2-4 hours depending upon conditions. Thorium extraction was reported to be similar to that in conventional atmospheric leaching. Other test data (9)(10) indicate that up to 95% sulphur extraction

can be achieved.

For this study, 90% sulphur removal has been accepted, leaving a sulphide sulphur content of approximately 0.3% S in the residual ore solids or 10% of the level with conventional leaching. A portion of the extracted sulphur will be converted directly to insoluble basic ferric sulphates. The balance will report as sulphuric acid and soluble iron sulphates which must be neutralized with limestone and lime. Limestone and lime requirements are higher than for the base case.

Thus, although the total sulphur content of the tailings solids is reduced by only the amount of reagent acid added to the conventional circuit, potential acid generation has been reduced to 1/10 of that in conventional tailings.

The only change in basic flowsheet is the substitution of autogenous pressure leaching for conventional atmospheric leaching.

Factors affecting the comparative economics of the process relative to the base case include:

- elimination of purchased acid and chlorate
- increased limestone-lime requirement
- much shorter leach time but more expensive vessels
- oxygen plant required
- reduction in steam requirement (none needed in leach)
- heat exchange (cooling) equipment required for leach discharge pulp
- gas handling and cooling system needed
- additional grinding costs (all testwork conducted at finer grind)
- possible increase in neutral and CCD thickener area and/or flocculant requirement

3.2.6 Radium Extraction from Tailings

Radium may be extracted from sulphuric acid leach tailings and subsequently

recovered from solution for separate disposal. Several extractants have been proposed for this purpose, including various salt solutions (11,12,15), EDTA (11,13,14,15,16) and DTPA (11).

Of various salt solutions tested, sodium chloride is preferred because of its low unit cost. Investigations have found that radium extraction kinetics with salt are fast, but that leaching is severely inhibited by low concentrations of sulphate ion (12). Ryan determined that multi-stage cross current contact with 3 M NaCl (after washing to remove sulphate) was capable of reducing radium levels to 2.3-11.4 Bq/g, up to an order of magnitude above the level considered necessary for unrestricted disposal.

Salt washing has not been selected for detailed cost assessment, and is considered an unattractive process for the following reasons:

- Assuming a solution strength of 3 M NaCl, in the order of 100 kg NaCl/t of ore must be discharged to the environment, unless facilities are provided for concentration of salt solutions (such as multiple-effect evaporation).
- Up to 99.9% removal of sulphate appears to be required to achieve a high degree of radium removal efficiency. In addition, sulphate remaining in the slurry will be largely transferred to the salt solution and must be removed before recycle. Ryan (12) proposes to remove sulphate by precipitation with barium (this would simultaneously remove radium); however, any excess barium must also be removed before recycle.
- High liquid/solid ratios (in the order of 10:1) appear to be required to effect substantial removal of radium.
- The process is not capable of producing waste that may be disposed of in an unrestricted manner, and the process has no effect on thorium.

Of the organic extractants, EDTA has been tested the most, and has been evaluated in detail. The basic process proposed by Nixon (13,14) has been chosen because it appears to be the furthest advanced, and sufficient information is available to allow construction of a conceptual flowsheet.

EDTA is effective in radium removal in the presence of sulphate, but the reagent also complexes other elements that are present, principally alkaline earths and iron. By washing sulphuric acid leached tailings to substantially reduce iron and calcium levels, Nixon was able to significantly reduce EDTA concentrations and consumption. The process may be described as follows:

> Conventional sulphuric acid leached tailings are washed for uranium recovery in a CCD circuit. Fresh water, rather than raffinate, is used as wash solution. The washed residue is contacted with EDTA in a CCD circuit, and radium-free residue is further washed in an additional CCD circuit for EDTA recovery. Radium may be recovered by ion exchange, but for purposes of this study it is assumed to be recovered by the less expensive route of direct precipitation with barium chloride. Complexed iron is subsequently removed by precipitation as the hydroxide, and EDTA is precipitated for recycle by acidification.

EDTA extraction of radium does not meet the objective of unrestricted residue disposal, and thorium distribution is not affected.

3.3 Hydrochloric Acid Leach

In the conventional sulphuric acid leach process essentially all of the radium and thorium (unless there is a thorium recovery circuit) reports to the solid tailings, along with unrecovered uranium.

CANMET have been investigating alternative processes for uranium ores in which the important radioactive elements (uranium, thorium, and radium) are extracted to yield tailings waste containing low levels of radioactivity, and with the radioacitve elements concentrated into forms suitable for long-term storage (17,18,19,21,22,23).

In contrast to the sulphate form, radium chloride is readily soluble in water and, in principle, radium can be efficiently extracted from the ore with hydrochloric acid. In practice, the presence of sulphate ion, originating

with the pyrite or other sulphides in the ore, has been found to interfere with radium dissolution. This interference is largely overcome by employing two stages of leaching wherein sulphate ion is essentially absent in the second stage. Radium extraction of about 95% can generally be obtained with the 2-stage technique.

Operating and capital costs for the process as applied to ore type A have been estimated and the process described by Lummus (1980)(22). Pregnant solution from the first stage leach is treated by solvent extraction to recover uranium, and thorium is recovered from the raffinate by a second solvent extraction procedure. Radium is removed from the final raffinate by co-precipitation with barium. The barium/radium precipitate is sent underground to a special disposal area. Thorium can be combined with the barium/radium for disposal or dried, packed in drums, and held for eventual sale.

The barren solution from radium removal is neutralized with limestone and lime to remove impurity metals, and sulphuric acid is then added to the solution to regenerate hydrochloric acid. After clarification, the weak hydrochloric acid is concentrated by distillation before being recycled to the leach step. The metal hydroxide precipitate and the gypsum sludge are directed to tailings.

Thorium solvent extraction could be eliminated and thorium permitted to leave with other impurity metals at the lime addition step. However, secure disposal of the rather voluminous sludge could well prove more costly than the solvent extraction procedure.

A major hurdle the flowsheet must overcome is to economically prevent discharge from the plant of large amounts of chlorides. The removal of chlorides from leached solids requires the addition of wash water, which reappears as pregnant solution leaving the leach circuit. The pregnant solution rate of 0.4 m^3/t employed in reference (22) would seem too low for adequate washing. An increase in this flow would increase the cost of hydrochloric acid regeneration.

The hydrochloric acid leach process requires that the leached ore have good filtration characteristics. No specific test information is available, but from what is otherwise known of these ores, type A would be expected to filter well, filtration of type B would be marginal, and type C ores, because of their clay content, would filter poorly.

Since acid run-off generated by oxidation of sulphide compounds is of concern with Ore A, the flowsheet includes a sulphide flotation step and subsequent roasting of the concentrate to make sulphuric acid.

3.4 Chlorination Process

Thorium and radium, and uranium in the presence of a reducing agent, may be converted to their chlorides by passing gaseous chlorine through a heated bed of ground ore (24,25). Much of the uranium and thorium chlorides volatilize and pass off with the exit gas; the radium chloride remains with the solids.

A significant difference from the hydrochloric acid leach is that only a weak solution of hydrochloric acid is required and regeneration of the acid is less costly.

A report prepared by the Minerals division of C.E. Lummus presented costs and a simplified flowsheet (22).

Ground and dried ore is contacted with chlorine in a fluid bed reactor; fluidization is assisted with nitrogen because oxygen must be excluded.

Depending on the composition of the ore, the reactor exit gas will contain chlorides of iron, nickel, and arsenic as well as those of uranium and thorium. With the addition of oxygen to the gas, the chlorides are converted to oxides. At the reaction temperature all of the oxides are solids with the exception of arsenic, and are recovered from the gas stream. Subsequently, the gas is cooled, and the condensed arsenic oxide is collected. Uranium is recovered from the oxide solids using a conventional carbonate leach process.

Chlorine is separated from the tail gas by water scrubbing, and recovered for reuse.

Radium and residual uranium and thorium are extracted from the chlorination calcine with hydrochloric acid. Reference (22) indicates this as a two-stage leach with conditions similar to those described in 3.3. However, laboratory work utilized a mild hydrochloric acid wash (25), and this technique has been accepted for this study.

Uranium is recovered from the leach liquor by solvent extraction, and radium removed from the raffinate with barium chloride. Hydrochloric acid is regenerated by sulphuric acid addition and concentrated by distillation; however with the dilute acid leach, the regeneration of the hydrochloric acid may not be necessary.

Thorium distributes between the carbonate leach residue, ore tailings, and leach solution, and is not collected for special disposal.

The presence of a reductant is required for the formation of uranium tetrachloride from ores by treatment with chlorine gas. In the case of type A ore, the reductant occurs naturally in the form of sulphides (S/U_3O_8 approximately 30/1). The sulphide content of ore types B and C is much lower (S/U_3O_8 approximately 0.25, and 0.5 respectively), and the addition of a reductant may be necessary for the reaction.

3.5 Carbonate Leach (ref. 26)

Carbonate, or alkaline leaching, is an established process for treatment of uranium ores, and is the conventional alternative to sulphuric acid leaching. Sodium carbonate is normally employed as the leaching agent, and the process operates at a pH of 9.5 to 10.0.

A number of carbonate leach mills have operated throughout the world, although

the only such Canadian plant was the Beaverlodge mill of Eldorado Nuclear Ltd. Carbonate leaching can normally compete economically with acid leaching only when acid consumption by the ore is high.

Ore is crushed and ground to a finer size than usually chosen for acid leaching. Grinding is conducted in process solution which contains about 50 g/L sodium carbonate as Na_2CO_3 and $NaHCO_3$, perhaps 50 g/L sodium sulphate, and some uranium. The ground product is thickened and leached. Air or oxygen may be used as an oxidant, and leaching is conducted at an elevated temperature. Sulphate and sulphides in the ore are largely or completely leached. Liquid/solid separation is preferably achieved by means of two or three stage filtration. Filtration is preferred over CCD because of reagent recovery considerations. Pregnant solution is clarified, may be partially evaporated and/or treated with lime to establish a solution and reagent balance, and the uranium precipitated with sodium hydroxide. The sodium diuranate formed does not meet refinery specifications, and must be dissolved and re-precipitated as a peroxide or ammonia product. Caustic barren solution from the initial yellowcake precipitation is recarbonated using boiler flue gas, and recycled to process. Barren solution from the final yellowcake precipitation must be discarded.

Because carbonate solutions are "valuable", the process is operated as essentially closed circuit, and water usage (and therefore mill effluent solution volume) is low relative to sulphuric acid leaching. As in sulphuric acid leaching, virtually all of the radium and thorium exit with the solid mill residue, although the radium is thought to be in the carbonate form.

Ore A contains uranium substantially in the form of the refractory mineral brannerite. Although no direct leach data are available on A type ores, experience with brannerite ores elsewhere indicate that uranium extractions would be unacceptably low. Therefore, application of carbonate leaching to Ore A is not considered to be a viable option.

B type ores are ammenable to carbonate leaching, and available metallurgical

data (5) indicate that uranium extractions can be equivalent to those obtained by acid leaching. Application of carbonate leaching to Ore B is considered in detail in Chapter 4.0.

There is some information related to carbonate leaching of C type ores (1), but unfortunately it is insufficient to allow a detailed analysis of the application and the environmental implications. Because of the significant sulphide sulphur level in the ore, reagent consumption in a carbonate circuit would be high (circa 35 kg/t). Economics suggest that flotation of sulphides followed by acid leach of concentration and carbonate leach of tailings would be required, but this "split" flowsheet is not particularly attractive because of the already small scale operation. In the absence of test data evaluation of such a flowsheet and prediction of the distribution and final form of the impurities of environmental concern is difficult.

Effluent solution is normally at a natural pH of about 10.0, and is accordingly low in most heavy metals. However, uranium in solution is high relative to acid leach effluents, and is a direct function of the efficiency of mill soluble uranium recovery.

3.6 Other Milling Alternatives

3.6.1 Heap Leaching

For purposes of this report, heap leaching includes dump leaching and heap leaching where ore is placed on a prepared surface pad for leaching, and the residue solids are not relocated for disposal.

Heap leaching using sulphuric acid has been applied commercially to a number of ores in the past. In general, it is considered only for low grade material that cannot be economically processed in a conventional mill, because uranium extractions are characteristically relatively low. Heap leaching may also prove to be economically attractive to higher grade ores that are metallurgically difficult to process in a conventional milling circuit.

Heap leaching is considered to be potentially applicable to Ore A which, because of ore grade, may not be able to support conventional processing. Ore A is expected to be amenable to heap leaching although quantitative information is limited. The use of bacterial leaching in a heap leach situation is attractive and would probably be employed.

The chemistry involved in heap leaching is not greatly different from conventional leaching in an environmental sense, but there are some process differences:

- The quantity of solution requiring neutralization and treatment for disposal will normally be greater.
- The heaps are underlain by an engineered base and a drainage system, which allows good control over seepage.
- Because the heaps are presumably above the water table, acid generation from pyrite oxidation can be expected. Also, because of the difficulty of placing an impermeable cap (with long-term integrity) over a large volume of tailings, long-term monitoring and treatment of seepage may be required. However, the volume of solution requiring treatment would be small.

From the above there is no fundamental environmental advantage or disadvantage to heap leaching, and no benefit is obtained relative to disposition of radionuclides or acid generating sulphides. Therefore, a detailed costing and environmental assessment of the process has not been undertaken in this report.

3.6.2 Nitric Acid Leach

Nitric acid has been proposed and tested as an alternative to sulphuric acid (11,15,27,28,29) and is capable of leaching radium, thorium and substantial amounts of sulphides. Ryan (27) reported up to 98% Ra removal and up to 99% Th removal from U.S. sandstone ores using a two stage, 3 M HNO_3 leach at 70°C. Skeaff (29) obtained 80% Th removal and perhaps 75-90% pyrite decomposition

from Elliot Lake ore under a variety of conditions with a free nitric acid level in excess of 10 g/L. In a later report (29) he notes that the radium extraction is only 5%. Other CANMET work has apparently not been formally reported. It is assumed that the difference in radium leaching behaviour between U.S. and Elliot Lake ores is related to the very aggressive conditions employed by Ryan in the U.S. case, and interference from sulphate in the Canadian case. Nitric acid leaching has not been demonstrated as a practicable method for radioisotope control with Canadian ores, and is therefore not examined in detail in this report.

Nitric acid has been proposed as a lixiviant for uranium and thorium, with radium subsequently removed by EDTA (29); the proposed process is extremely complex. The benefits of nitric acid in this case include only removal of sulphide sulphur and an increase in thorium extraction from 70% to 80%. The marginal improvement in thorium removal is not attractive in itself, and it is concluded that, based on reported testwork, the incentive for use of nitric acid lies primarily with sulphide removal and subsequent reduction of tailings acid generating capacity.

When assessed from the point of view of sulphide sulphur oxidation, nitric acid leaching is best compared with a pressure sulphide oxidation process. The reduction of nitric acid produces primarily nitric oxide gas. In principle, the acid can be regenerated externally to the leach circuit, or internally using oxygen. In the limit, the process reduces to a nitric acid catalysed sulphuric acid pressure leach. The choice between such a process and a "conventional" oxygen pressure leach is primarily one of economics, but the nitric acid alternative has the environmental disadvantage of contributing nitrate to effluent streams.

Both the environmental implications, and the process operating costs for nitric acid leaching depend on efficient recycle of nitrate. Skeaff (29) reports that at 60-75°C, and a 30 hour leach time, the nitrate content of the leach liquor must be in the order of 50 g/L or more for effective pyrite oxidation. The proposed HNO_3-EDTA process (29), which employs external

regeneration of nitric acid, shows a residual nitrate level in leach solution of about 2 g/L (and about 86% pyrite oxidation). These latter values do not appear to be compatible with the reported testwork, and are probably unattainable.

3.6.3 Preconcentration

Considerable testwork has been conducted on preconcentration of uranium ores, particularly by CANMET and operators in the Elliot Lake (Ore A) area. The primary processes tested are flotation, heavy media separation, radiometric sorting, and high intensity wet magnetic separation. Until recently, all testwork has been directed toward separation of uranium only into a relatively small fraction of the ore. To our knowledge, and with the exception of certain Canadian ores not addressed in this study, all such attempts have failed in an economic sense, mainly because of the cost penalty associated with loss of a fraction of the uranium to preconcentration tailings.

More recently, CANMET (3) tested flotation and magnetic separation processes for Ore A on laboratory and pilot scale, and the economics were briefly evaluated by Lummus(30). The purpose of these studies was to separate pyrite and radionuclides into a small fraction of the ore, and to produce a preconcentration tailings suitable for more-or-less unrestricted discharge to the environment. Lummus concluded that the proposed preconcentration process would not be economically attractive for a sulphuric acid leach plant, but would be for a hydrochloric acid process. However, the cost penalty for the sulphuric acid leach case inferred from Lummus data is modest compared to other processes considered in this study and preconcentration could be attractive if it led to a significant improvement in an environmental sense.

The CANMET study showed that 93% of the uranium, 90.5% of the thorium, and 87.5% of the radium could be concentrated in 26% of the ore weight, with the preconcentration tailings comprising 71% of the ore weight, with a radium content of 17 Bq/g. While these results are considered remarkably good in general, the quantity of concentrate is sufficiently high that it must be

disposed of in a conventional surface tailings pond. The creation of two, rather than one, tailings disposal facilities (both of a substantial size) does not improve the environmental situation, even though one of the ponds may not require continued monitoring. In this respect, preconcentration is somewhat similar to pyrite and radionuclide flotation from tailings, discussed in Sections 3.2.2 and 3.2.3.

It is concluded that unless the ratio of concentration can be substantially improved so that alternate disposal means for the concentrate residue can be considered (such as underground disposal), preconcentration does not materially improve the environmental situation.

3.6.4 In-Situ Leaching

In-situ leaching has been employed for uranium recovery, particularly from U.S. sandstone ores. Although it is difficult to compare the degree of uranium extraction by in-situ leaching with that obtained by conventional mining and milling, it will in general be lower, and the method is normally applied to ore-bodies that cannot be exploited economically by conventional means.

To be amenable to in-situ leaching, an ore-body must be adequately permeable, below the water table, and underlain by impermeable material. Alternately, the ore may be broken and leached in place, if it is surrounded by competent rock. Although each of the three ore types present significant difficulties for in-situ leaching, it is possible that techniques can be developed that are economically applicable to specific situations.

In-situ leaching is environmentally different from conventional processing. Most or all of the ore remains in place, along with the bulk of the environmentally significant elements. In general, however, a certain amount of groundwater must enter the system in order to ensure containment of leach solutions, and this excess must be treated on surface for removal of radium and other contaminants before release to the environment.

During operation of an in-situ project, water management ensures that leach solutions underground are controlled and not lost to the surrounding environment. Restoration of the depleted ore zone or associated aquifer may not be easily accomplished, depending on the leach system employed, and the environmental guidelines imposed. For example, U.S. in-situ operations employing ammonia based reagents have found it difficult to restore aquifers to the required background ammonia levels.

The detailed assessment of in-situ leaching for the three cases under study is considered beyond the scope of this project, but the applicability of this technique to specific defined ore bodies may be of some environmental interest.

3.6.5 Ferric Chloride and Ferric Sulphate Leaching

From an environmental viewpoint, ferric chloride and ferric sulphate leaching of uranium are not far removed from hydrochloric acid and sulphuric acid leaching respectively. They have therefore not been separately studied.

3.6.6 Caro's Acid and Persulphuric Acid Leaching

These reagents are essentially alternate choices of oxidant and have no environmental significance except that they both can substitute for sodium chlorate and thereby reduce the chloride level in effluents. The choice of either of these reagents is considered one of economics only.

3.7 Selected Process Alternatives

Based upon the review presented in this chapter several milling alternatives were considered worthy of further evaluation for their cost, health and safety concerns and environmental aspects. Base Case process selections are based upon economics while alternatives provide technically viable process options which produce environmentally more acceptable tailings. Although more

acceptable over the long term, their operational impacts may be greater.

The processes selected are as follows:

i) Ore A

Base Case - Sulphuric Acid Leaching
Alternative 1 - Pyrite Flotation
Alternative 2 - Acid Production
Alternative 3 - Bulk Flotation
Alternative 4 - Thorium Removal
Alternative 5 - Autogenous Leach
Alternative 6 - EDTA Leach
Alternative 7 - HCl Leach

ii) Ore B

Base Case - Sulphuric Acid Leaching
Alternative 1 - HCl leach
Alternative 2 - HCl + Cl_2 Leach
Alternative 3 - Carbonate Leach

iii) Ore C

Base Case - Sulphuric Acid Leaching
Alternative 1 - HCl Leach
Alternative 2 - HCl + Cl_2 Leach

A process flowsheet and cost estimate is developed for each of the above processes in the following chapter.

4.0 DEVELOPMENT OF PROCESS FLOWSHEETS AND COST ESTIMATES

4.1 General

4.1.1 Design Capacity

Process flowsheets were developed for the base case processing options for the three ore types. These flowsheets are based upon 20 years of production at 2.26×10^6 kg U_3O_8/a. Plant availability and percent recovery have been assumed based upon operating experience, resulting in the design capacities for mill production as noted for each alternative.

Because the front-end grinding circuits are common to many alternatives, for cost comparisons the mill capacity has been retained constant. Therefore with some alternatives, the annual production of U_3O_8 may be marginally greater or lower than that for the base case. The incremental difference in recoveries is handled in the economic comparison.

4.1.2 Costing Data

Data sources included:

- detailed design and construction estimates for existing facilities
- general cost estimates reported in literature
- supplier quotations
- general cost data on file

Where necessary, all costs were normalized to mid-1985 by adjusting the data with standard engineering construction cost indices.

Except where noted, equipment prices were individually obtained from telephone quotations or estimated from A.H. Ross & Associates cost data. Installation costs and other direct costs were obtained by factoring equipment costs based on A.H. Ross & Associates cost data for similar installations. Reagent costs are based primarily on vendor telephone quotations.

A contingency allowance of 20% has been applied to all capital costs, and an allowance of 10% has been provided for operating costs. All capital costs are amortized over 20 years at 12%. This is not the approach used for a true economic evaluation as amortization periods of 2 to 5 years would be typical. The 20 year period was selected to permit a better comparison of all aspects of the total production costs.

4.1.3 Capital Costs

The capital costing data include:

- all site preparation costs, including general yard facilities related to the process plant such as power and water distribution, fire protection, sewers, monitoring ponds, etc.
- process plant and buildings
- laboratory and mill offices
- tailings treatment facilities and tailings lines
- tailings pond and effluent treatment ponds
- construction indirect costs
- engineering, procurement and construction management
- contingency allowance

Capital costs exclude:

- ancillary facilities (offices, warehousing, shops)
- mine costs
- offsite facilities and supply of services to the site
- escalation and interest costs
- working capital and other Owners costs

4.1.4 Operating Costs

Operating costs include only direct mill and effluent treatment costs, and exclude:

- mine costs
- administration (other than direct mill administration)
- security and safety personnel
- head office costs
- local taxes, insurance, licences, etc...

Labour rates have been drawn from several sources and are appropriate for the chosen locations. A payroll burden of 30% has been applied to these rates.

Maintenance costs are estimated as a percentage of capital costs. Power costs are based on hydro power for Ore A, and diesel power for Ores B and C. Fuel for steam generation is oil except for those cases where consumption is high enough to justify coal.

4.1.5 Process Flowsheets

A process flowsheet and description was prepared for each alternative. For each flowsheet, design criteria are clearly summarized for the major unit operations (i.e. crushing, grinding, leaching, liquid-solid separation, product concentration, uranium recovery and tailings neutralization). A capital and operating cost summary highlighting the expenditures for the major unit operations and physical support facilities is presented in this section. Details of equipment requirements and costing for both capital and operating items is provided in Appendix A.

4.2 Process Description and Cost Estimates for Ore A Milling Alternatives

4.2.1 Base Case - Sulphuric Acid Leach

i) Process Description and Flowsheet

The conventional treatment for type A ore (Elliot Lake) is the atmospheric

sulphuric acid leach. Specific flowsheet details are as follows:

The ore is crushed underground to minus 200 mm and is transported to surface storage silos. Pebbles for use as grinding media are separated by screening. The ore is ground to 80% passing 65 mesh in a semi-autogenous primary mill and secondary pebble mill using cyclones for classification. Ground ore is thickened and filtered. Filter cake is repulped with recycled pregnant solution to a density of 76% solids entering leach.

Leaching is performed in two lines of air agitated pachucas providing 43 hours retention time. Sodium chlorate is added as oxidant and the pulp temperature is raised by introduction of live steam. Uranium extraction is 95% and sulphuric acid is purchased.

Leach discharge pulp is separated into slimes and sands for separate washing in a 5-stage CCD thickening circuit and a 4-stage cyclone system respectively. The combined soluble loss from the solids washing circuits is 0.7% of the uranium in the ore. The volume of uranium-bearing pregnant liquor is 2.25 m^3/t ore.

The pregnant solution is clarified in a hopper clarifier and sand filters. The uranium is then recovered from the solution by solvent extraction using a tertiary amine in kerosene diluent.

The uranium is stripped from the solvent by ammonia and is also precipitated with ammonia. Yellowcake is washed in a thickener and a centrifuge and is then dried in an oil-fired multi-hearth furnace.

Ammonium sulphate is recovered by crystallization from a bleed stream of barren strip solution.

Solvent extraction raffinate is recycled as wash for leached solids.

The washed sands and slimes are combined and neutralized with limestone and lime in air-agitated pachucas before discharge to the tailings pond. Pond overflow solution is treated with barium chloride to precipitate the radium.

A block flowsheet is presented on Figure 4.2.1.

ii) Selected Design Criteria

Crushing	- underground to 200 mm
Grinding	- 80% passing 65 mesh
Leaching	
Retention time	- 43 hours
Feed density	- 76% solids
Temperature	- 77°C
Sulphuric acid (100%)	- 33.5 kg/t
Sodium chlorate	- 1.5 kg/t
Uranium extraction	- 95%
Liquid-Solids Separation	
Cyclone stages (sands)	- 4
CCD stages (slimes)	- 5
soluble loss	- 0.7%
Solvent Extraction	
Pregnant solution	- 2.25 m^3/t ore
Amine strength	- 2.5 v/v%
Stages - extraction	- 4
- scrub	- 1
- strip	- 4
- regeneration	- 1
Ammonia	- 0.4 kg/kg U_3O_8
Uranium Recovery	
Ammonia	- 0.2 kg/kg U_3O_8
Ammonium sulphate concentration	- 150 g/L

Tailings Neutralization

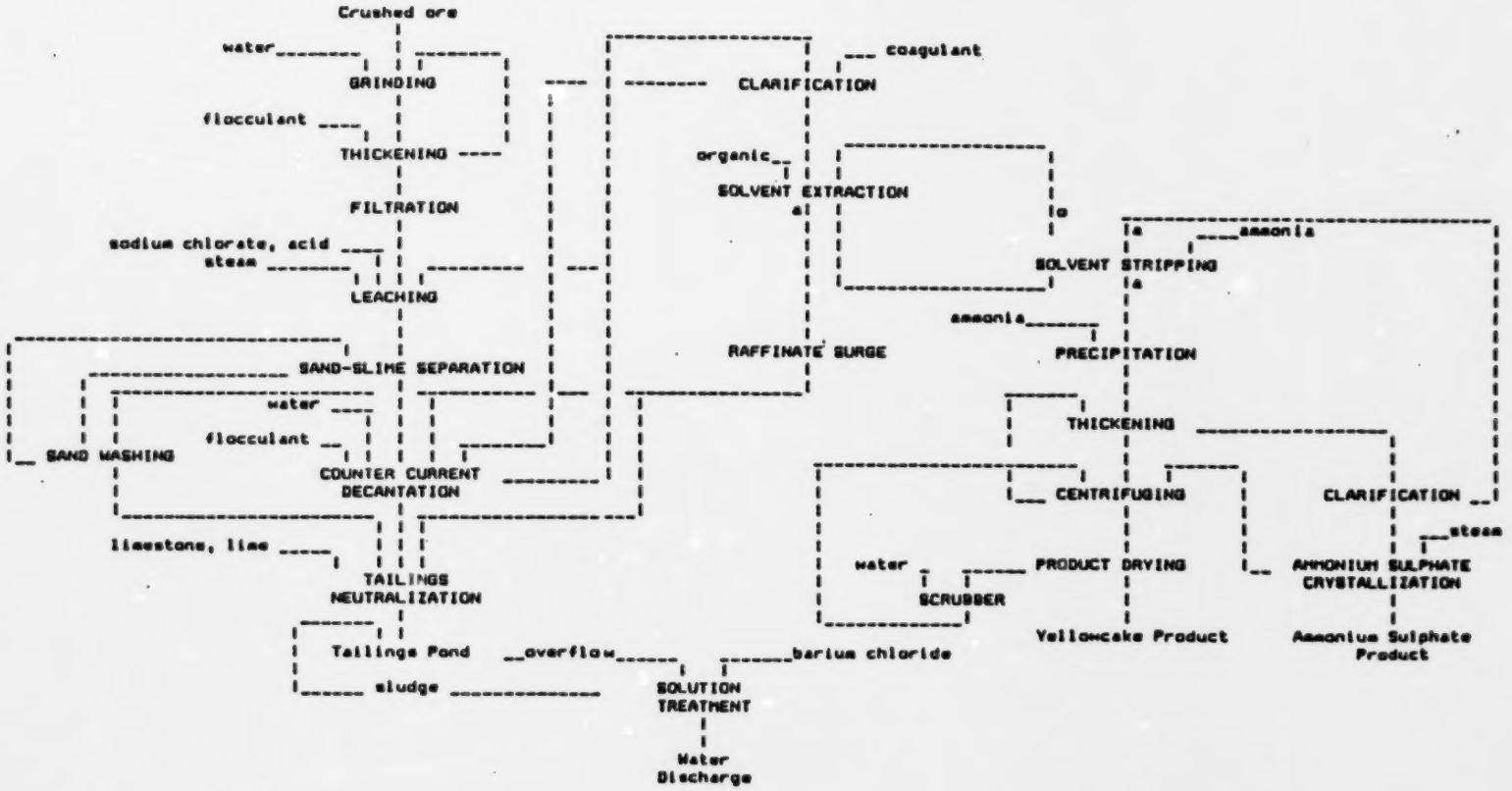
Limestone	- 18 kg/t
Lime	- 11 kg/t

iii) Capital Costs

Summary of Capital Costs	$
Ore receiving and storage	7,050,000
Grinding and thickening	12,690,000
Neutral filtering	4,660,000
Leaching	9,140,000
Sands washing - CCD thickening	8,760,000
Clarification	3,740,000
Solvent extraction	5,950,000
Uranium recovery	3,330,000
Tailings neutralization	2,860,000
Ammonium sulphate recovery	1,940,000
Reagent handling	4,830,000
Utilities	5,710,000
Buildings	17,820,000
Total direct cost	88,480,000
Indirect costs	26,540,000
Mobile equipment	60,000
Total before contingency	115,080,000
Contingency @20%	23,020,000
Total Estimated Capital Cost	138,100,000
Amortized Cost (20 yrs at 12%)	$17,910,000/a
$/t ore	$6.73
$/kg U_3O_8	$7.93

Figure 4.2.1

BASE CASE SULPHURIC ACID LEACH



A.H. ROSS & ASSOCIATES

iv) Operating Costs

Summary of Operating Costs	$
Supervision	949,000
Operating Labour	2,827,000
Maintenance	4,634,000
Process chemicals	9,803,000
Operating supplies	1,321,000
Power	2,998,000
Fuel oil	7,141,000
Water	1,118,000
Analytical and test laboratories	626,000
Sub-total	$31,417,000
Contingency @10%	3,142,000
Total operating cost	34,559,000
$/t ore	$12.99
$/kg U_3O_8	$15.29

4.2.2 Alternative 1 - Pyrite Flotation

i) Process Description and Flowsheet

This alternative involves an add-on to the base case sulphuric acid leach flowsheet. The primary objective is to reduce the quantity of potential acid generating material (pyrite) discharged to the main tailings disposal area. However, the pyrite concentrate can be used to manufacture acid for internal consumption and for sale.

Washed, but unneutralized or only partially neutralized, leach residue is treated by standard xanthate flotation to produce a pyrite concentrate. The rougher concentrate is reground and refloated in two stages of cleaners. The cleaner concentrate pulp is thickened and pumped to a disposal area separate from the main plant tailings ponds.

Flotation tailings are pumped to the existing base plant neutralization circuit.

The cleaner concentrate is approximately 6% of the ore weight and contains 90% of the sulphide sulphur in the ore feed. It is assumed that the concentrate will be disposed in specially designated abandoned stopes of the underground workings. Alternatively special on-surface disposal areas could be considered. These costs are included as part of the waste management costs addressed in Chapter 6.0. Neither option is likely to be greeted with enthusiasm by a mine operator.

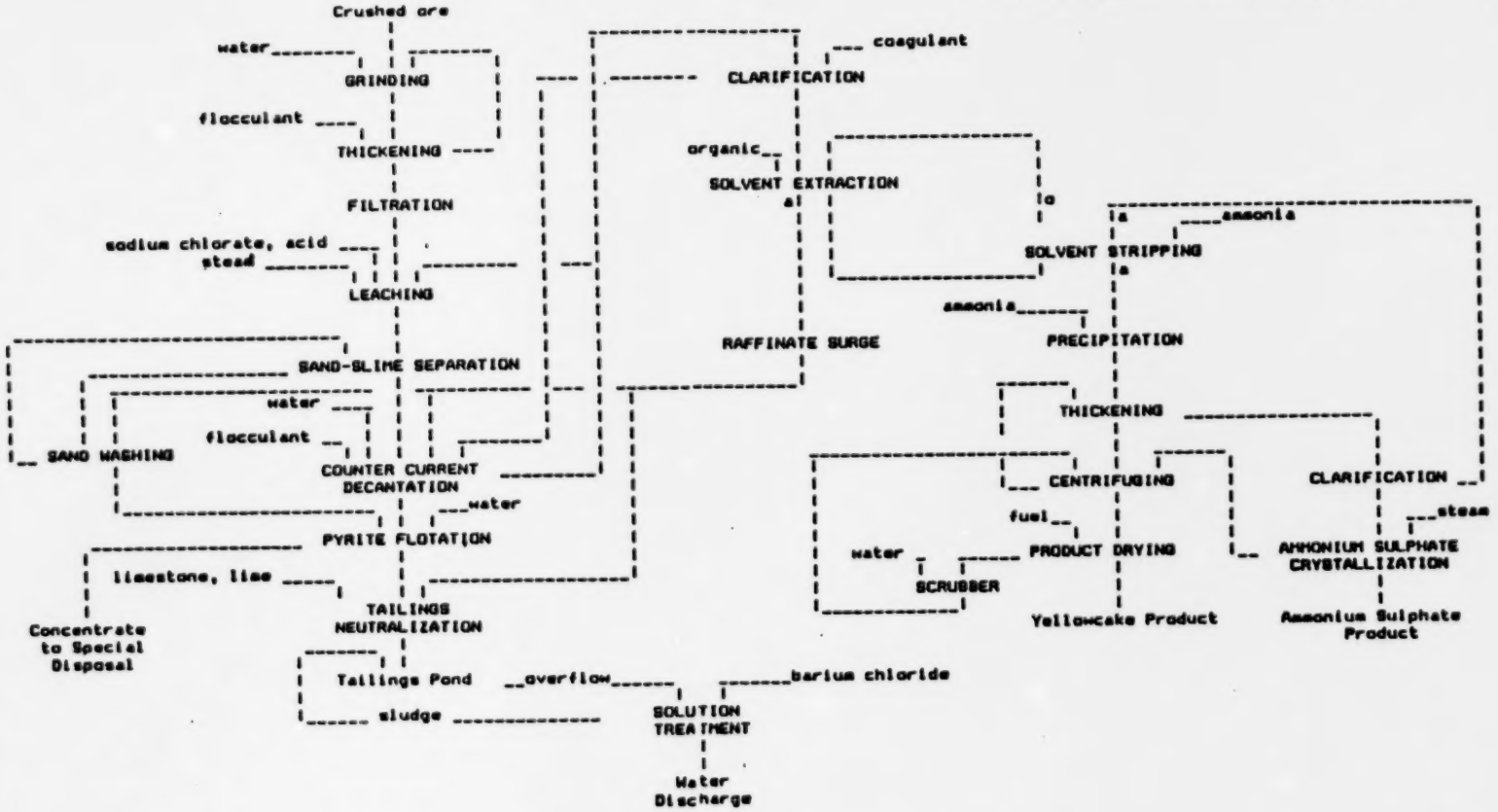
A block flow diagram is provided on Figure 4.2.2.

ii) Selected Design Criteria

Rougher Flotation	
Pulp density	- 40% solids
pH	- 3-4
Retention time	- 15 min.
Regrind	- 80% passing 325 mesh
Cleaner flotation	
Pulp density	- 30% solids
Stages	- 2
Retention time/stage	- 12 min.

Figure 4.2.2

ALTERNATIVE 1 - PYRITE FLOTATION



Flotation Reagents

Potassium amyl xanthate	- 0.10 kg/t
Pine oil	- 0.65 kg/t

Pyrite concentrate

Grade	- 48% S
Weight	- 5.6%
Recovery of sulphide S	- 90% or ore

Pyrite Tailings

Grade	- 0.3% S
Residual sulphide S	- 10% of ore

iii) Capital Costs

Summary of Capital Costs	$
Incremental cost of flotation circuit	
Direct cost of facilities	5,920,000
Indirect costs	1,770,000
Total before contingency	7,690,000
Contingency @20%	1,540,000
Total Cost	9,230,000
Cost for Base Case	138,100,000
Total estimated capital cost	147,330,000
Amortized cost (20 yrs at 12%)	$19,110,000/a
$/t ore	$7.18
$/kg U_3O_8	$8.46

iv) Operating Costs

Incremental Costs	$
Supervision	52,000
Operating labour	341,000
Maintenance	323,000
Process chemicals	1,534,000
Operating supplies	290,000
Power	341,000
Fuel oil	-
Water	-
Analytical and test laboratories	58,000
Sub-total	2,939,000
Contingency @10%	294,000
Total incremental cost	3,233,000
Cost for base case	34,559,000
Total operating cost	$37,792,000
$/t ore	14.27
$/kg U_3O_8	16.72

4.2.3 Alternative 2 - Acid Production

i) Process Description and Flowsheet

This alternative is a simple add-on to the Alternative 1 - Pyrite flotation. In this option, rather than dispose of the concentrate, the pyrite concentrate is roasted and sulphuric acid is manufactured from the roaster gases in a conventional contact acid plant. Roaster calcine is quenched and pumped to the main plant tailings disposal. In this option a filtering stage is provided to further dewater thickened cleaner concentrate for feeding the roaster.

The block flow sheet is shown on Figure 4.2.3.

ii) Selected Design Criteria

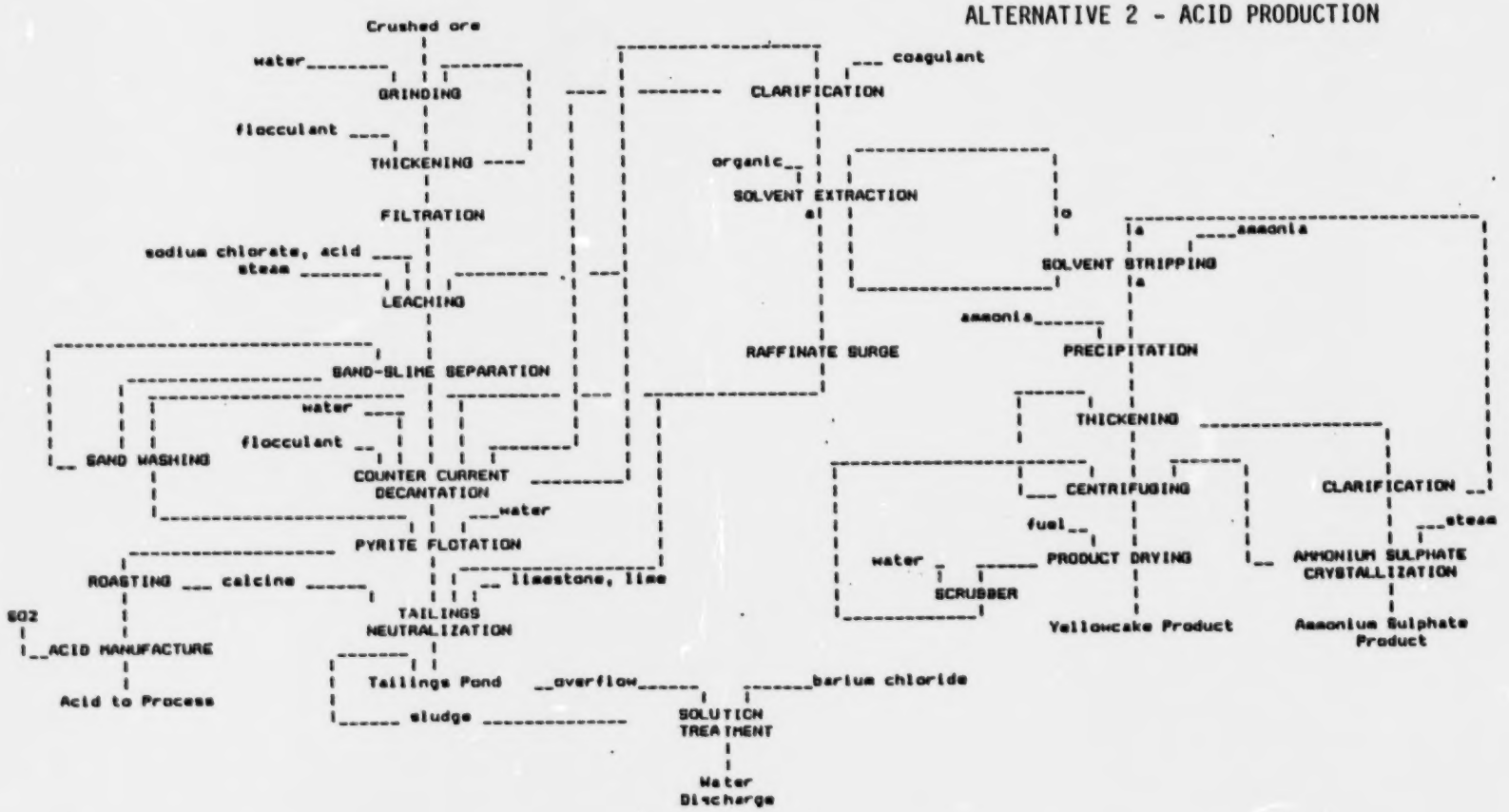
Pyrite Flotation	- as for Alternative 1
Roaster Feed	- wet filter cake
Acid Plant	
Acid production (100% basis)	- 210, 180 t/a
Sulphur recovery to acid	- 95.5% of S in concentrates
Sulphur to stack	- 0.7% of S in roaster gas

iii) Capital Costs

Summary of Capital Costs	$
Incremental cost of flotation circuit and acid manufacture facilities.	
Total installed cost including contingency	
Pyrite flotation, filtration	10,880,000
Roaster facility	13,620,000
Acid manufacturing plant	25,940,000
Total cost	$50,440,000
Cost for base case	138,100,000
Total estimated capital cost	188,540,000
Amortized Cost (20 yrs at 12%)	$24,450,000/a
$/t ore	$9.91
$/kg U_3O_8	$10.82

Figure 4.2.3

ALTERNATIVE 2 - ACID PRODUCTION



A.H. ROSE & ASSOCIATES

iv) Operating costs

Incremental Costs	$
Supervision	266,000
Operating labour	683,000
Maintenance	2,359,000
Process chemicals	1,534,000
Operating supplies	290,000
Power	1,282,000
Fuel oil	55,000
Water	99,000
Analytical and test laboratories	111,000
	6,679,000
Less: Credit for acid produced and used in uranium leaching (1)	2,678,000
Steam credit	2,695,000
Sub-total	$1,306,000
Contingency @ $10%	131,000
Total incremental cost	$1,437,000
Cost for base case	34,559,000
Total operating cost	$35,996,000
$/t ore	$13.53
$/kg U_3O_8	$15.93

(1) no credit or penalty taken for acid production in excess of process requirements. Excess amounts to 120,920 t/a (100% basis)

4.2.4 Alternative 3 - Radioisotope Bulk Flotation

i) Process Description and Flowsheet

This alternative is a supplementary add-on to Alternative 1 pyrite flotation. The objective is to greatly reduce the quantity of radioactive material (radium, thorium and uranium) in the main tailings disposal area and to produce a reduced volume of radioactive tailings for special disposal. Some additional pyrite is also removed. This procedure can be included with either the pyrite flotation or the acid production option.

Pyrite flotation tailings are subjected to flotation treatment using oleic acid as a collector. The flotation concentrate pulp is thickened and assumed to be pumped to predesignated underground stopes. The flotation tailings are returned to the leach plant neutralization circuit.

A block flowsheet is present on Figure 4.2.4

ii) Selected Design Criteria

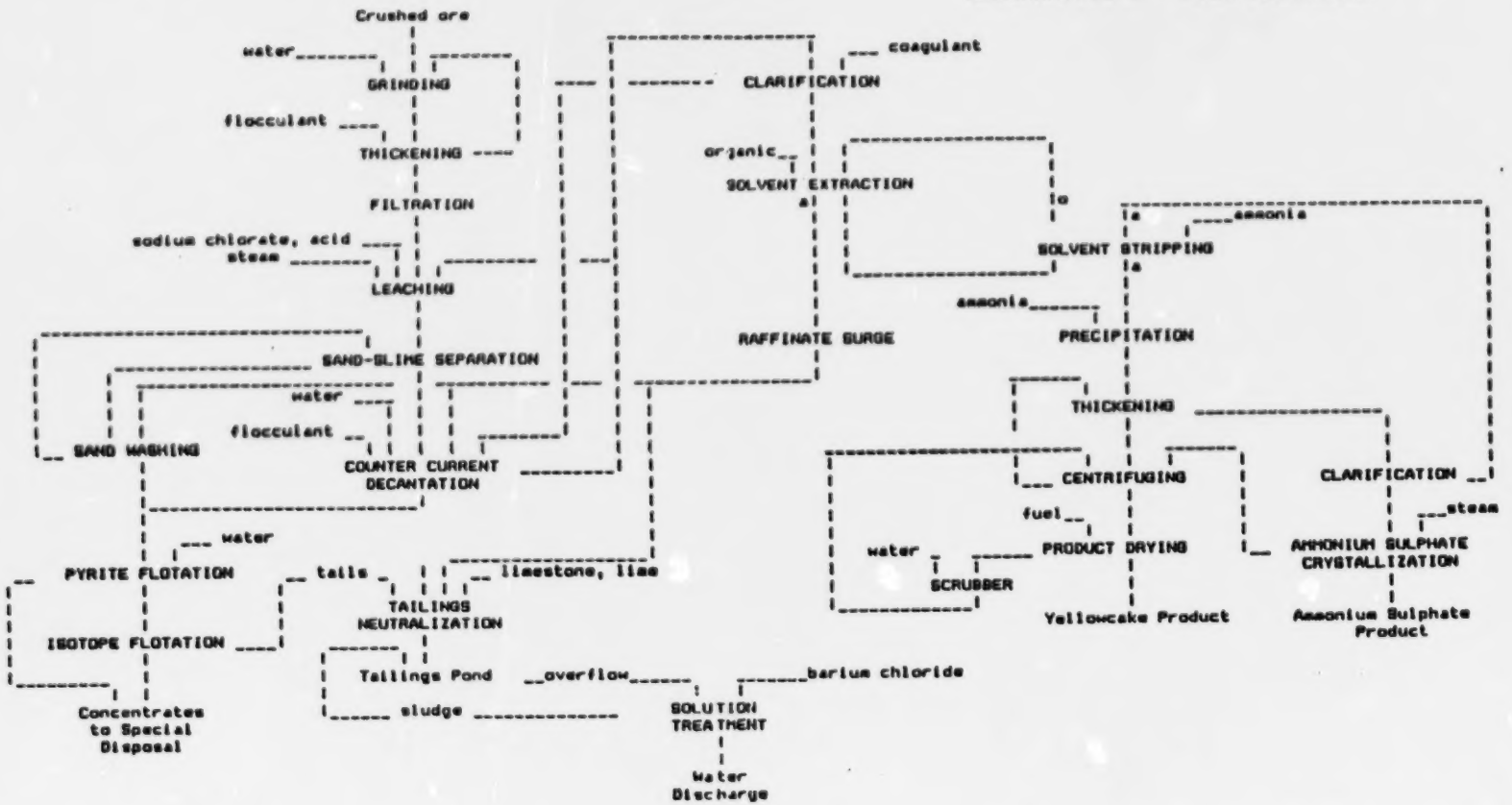
Flotation

Pulp density	- 35% solids
Retention time	- 30 min.
Oleic acid	- 1.55 kg/t

Radioisotope Concentration	Approximate Composition	% of Ore
Weight		24.4
S	1.0%	8
U_3O_8	0.009%	2.4
Thorium	0.06% (0.02%)	62(19)
Th-230	23.6 Bq/g (2.2 Bq/g)	62(6)
Ra-226	23 Bq/g	60

Figure 4.2.4

ALTERNATIVE 3 - BULK FLOTATION



A.H. ROSS & ASSOCIATES

Flotation Tailings	Approximate Composition	% of Ore
Weight		70
S	0.1%	2
U_3O_8	0.003%	2.5
Thorium	0.01%(0.01%)	35(22)
Th-230	4.3 Bq/g (0.7 Bq/g)	35(6)
Ra-226	4.0 Bq/g	35

Figures in brackets refer to plant which includes a circuit to recover thorium from uranium solvent extraction raffinate.

iii) Capital Costs

Summary of Capital Costs	$
Incremental cost of radioisotope flotation	
Direct cost of facilities	3,330,000
Indirect costs	1,000,000
Total before contingency	4,330,000
Contingency @20%	860,000
Total cost	5,190,000
Cost for base case	138,100,000
Cost of pyrite flotation	9,230,000
Total estimated capital cost	152,520,000
Amortized Cost (20 yrs at 12%)	$19,780,000/a
$/ore	$7.43
$/kg U_3O_8	$8.75

iv) Operating Costs

Incremental Costs	$
Supervision	-
Operating labour	171,000
Maintenance	181,000
Process chemicals	4,380,000
Operating supplies	-
Power	202,000
Fuel oil	-
Water	-
Analytical and test laboratories	-
Sub-total	4,934,000
Contingency @10%	493,000
Total incremental cost	5,427,000
Cost for base case	34,559,000
Cost of pyrite flotation	3,233,000
Total operating cost	43,219,000
$/t ore	$16.25
$/kg U_3O_8	$19.12

4.2.5 Alternative 4 - Thorium Recovery

i) Process Description and Flowsheet

This alternative is an add-on to the base case process. Uranium plant raffinate is advanced to a second solvent extraction circuit for recovery of the contained thorium. No pH adjustment or pretreatment is required. The solvent is a mixture of alkylphosphoric acid (DEHPA) and tributyl phosphate (TBP) in kerosene. The thorium is stripped from the solvent by moderately strong sulphuric acid solution and immediately precipitates as thorium sulphate. The precipitate is thickened and filtered and the wet filter cake is

rummed for special disposal or for sale as a byproduct. For this assessment special disposal has been assumed.

Approximately 10% of the stripped solvent is scrubbed with hydrofluoric acid before recycle to the extraction stage.

Thorium plant raffinate is returned to the uranium plant for washing leached olids. A block flow diagram is provided on Figure 4.2.5.

i) Selected Design Criteria

olvent extraction

Feed solution (uranium raffinate)	- 0.07 g/L Th
Stages - extraction	- 3
- strip	- 2
- scrub	- 1
Solvent	- DEHPA/TBP in kerosene
Strip solution	- 20 N sulphuric acid

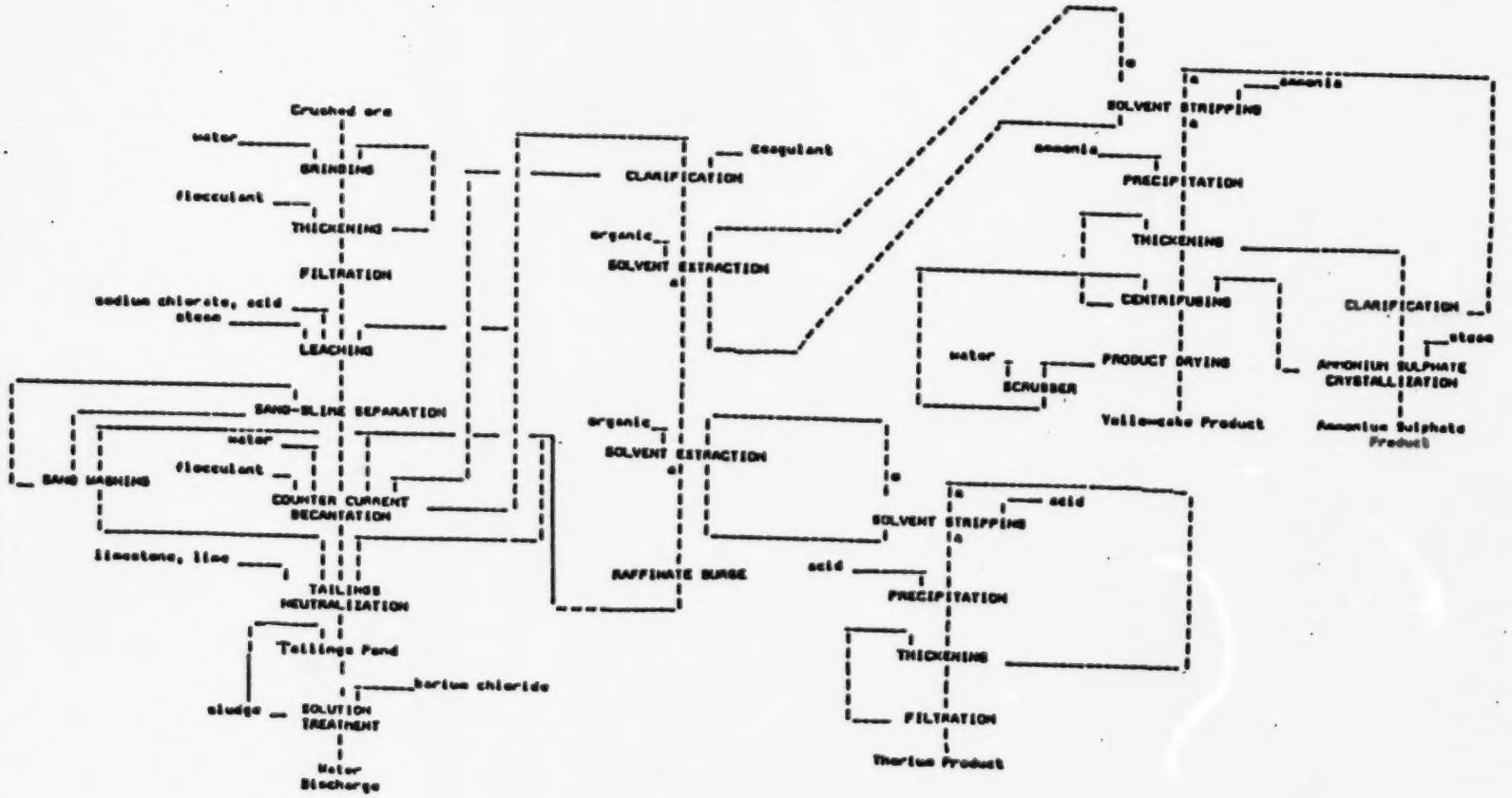
roduct

Weight	- 2330 kg/day
Grade	- 44% Th
	- 26,000 Bq/g Th230
Recovery	- 94% of Th in feed solution
	- 96% of Th in ore
	- 88% of Th-230 in ore

iii) Capital Costs

Summary of Capital Costs	$
Incremental cost of thorium circuit	
Solvent extraction	5,950,000
Dewatering, drumming	1,790,000

ALTERNATIVE 4 - THORIUM PRECIPITATION



A.H. ROSE & ASSOCIATES

Building	1,940,000
Total direct costs	9,680,000
Indirect costs	2,910,000
Total before contingency	12,590,000
Contingency @ 20%	2,520,000
Total Cost	$15,110,000
Cost for base case	138,100,000
Total capital cost	153,210,000
Amortized cost (20 yrs at 12%)	$19,870,000/a
$/t ore	$7.47
$/kg U_3O_8	$8.79

iv) Operating Costs

Incremental Costs	$
Supervision	-
Operating labour	228,000
Maintenance	529,000
Process chemicals	2,178,000
Operating supplies	136,000
Power	71,000
Fuel Oil	250,000
Water	10,000
Analytical and test laboratories	51,000
Sub-total	3,453,000
Contingency @10%	345,000
Total Incremental Cost	3,798,000
Cost for Base Case	34,559,000
Total operating cost	38,357,000
$/t ore	$14.42
$/kg U_3O_8	$16.97

4.2.6 Alternative 5 - Autogenous Leach

i) Process Description and Flowsheet

The flowsheet is basically the same as for the base case. Only the differences are outlined in this section.

Additional grinding capacity is provided to grind the ore to 80% passing 150 mesh as all testwork has been preformed at this finer grind.

Neutral filter duty is somewhat reduced because the pressure leaching is conducted at a lower pulp density (60% solids) than in conventional leaching practice, and acid, chlorate, and steam additions are eliminated. The volume of recycled pregnant solution for repulping the neutral filter cake has been increased to maintain the same volume of solvent extraction feed solution as in the base case.

Leaching is performed in 3 lines of 4-compartment horizontal autoclaves providing 4 hours retention time; there are four autoclaves in each line. Oxygen, produced in an on-site package oxygen plant, is introduced into the pulp to promote oxidation of the sulphide sulphur. The reaction generates more than sufficient heat to maintain the required 160°C leach temperature, and a gas cooling system is provided to dissipate the excess. Autoclave pressure is 1138 kPa absolute.

Leach discharge pulp is cooled to approximately 60°C in a double-pipe heat exchanger using water as the cooling media. The cooled pulp is advanced to sand-slime separation and washing.

Because of the finer grind, there is a greater quantity of slimes than in the base case and larger CCD thickeners may be required.

However, there is no specific data in this regard, and for the purpose of this

study, the same solids washing equipment has been provided as for the base case.

The solvent extraction and uranium recovery circuits are identical to those for the base case.

A water cooling tower is provided for treating discharge water from the pulp cooling heat exchangers and from the gas cooling vessels.

Figure 4.2.6 is a block flowsheet for the pressure leaching circuit.

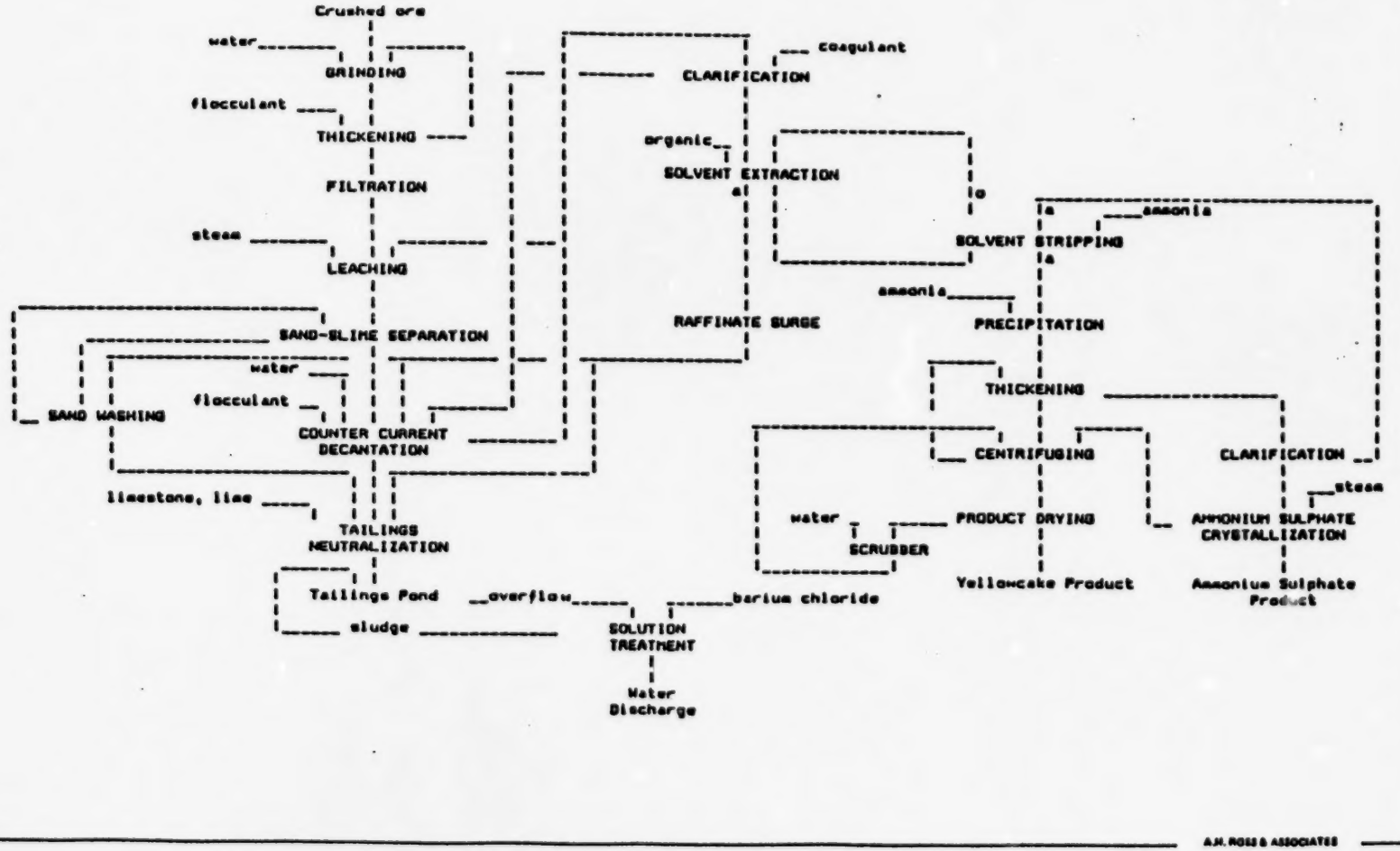
Acid generation is estimated to be approximately 37 kg/t of ore (100% basis), or just slightly more than the 33.5 kg/t of acid requirement for the base case.

ii) Selected Design Criteria

Crushing	- underground to 200 mm
Grinding	- 80% passing 150 mesh
Leaching	
Retention time	- 4 hours
Feed density	- 60% solids
Temperature	- 160°C
Pressure	- 1138 kPa (absolute)
Oxygen	- 54.7 kg/t
Uranium extraction	- 95%
Sulphide sulphur oxidation	- 90%
Tailings Treatment	
Limestone	- 36.9 kg/t
Lime	- 18.5 kg/t

Figure 4.2.6

ALTERNATIVE 5 - AUTOGENOUS LEACH



iii) Capital Costs

Summary of Capital Costs	$
Ore receiving	7,050,000
Grinding and thickening	16,130,000
Neutral filtering	3,660,000
Leaching	37,210,000
Sands washing- CCD thickening	8,760,000
Clarification	3,740,000
Solvent extraction	5,950,000
Uranium recovery	3,330,000
Tailings neutralization	2,860,000
Ammonium sulphate recovery	1,940,000
Reagent handling	4,390,000
Utilities	3,040,000
Buildings	17,870,000
Total direct cost	115,930,000
Indirect costs	34,780,000
Mobile equipment	60,000
Total before contingency	150,770,000
Contingency @20%	30,150,000
Total estimated capital cost	180,920,000
Amortized Cost (20 yrs at 12%)	$23,470,000/a
$/t ore	$8.82
$/kg U_3O_8	$10.38

iv) Operating Costs

Summary of Operating Costs	$
Supervision	1,009,000
Operating labour	2,998,000

Maintenance	6,333,000
Process chemicals	7,150,000
Rental on oxygen plant	5,088,000
Operating supplies	1,321,000
Power	3,937,000
Fuel oil	3,551,000
Water	1,073,000
Analytical and test laboratories	626,000
Sub-total	33,086,000
Contingency @10%	3,309,000
Total operating cost	36,395,000
$/t ore	$13.68
$/kg U_3O_8	$16.10

4.2.7 Alternative 6 - EDTA Leach

i) Process Description and Flowsheet

The Base Case process is essentially unchanged for EDTA recovery of radium, except that raffinate is no longer recycled to CCD.

Feed to the EDTA circuit is washed tailings residue from the last uranium CCD thickener. Washing in the uranium CCD circuit is accomplished with barren acid solution from EDTA precipitation rather than uranium circuit raffinate, because it is necessary to substantially remove calcium and iron from the slurry.

The EDTA circuit feed is slightly acidic and is neutralized with caustic. Subsequently, radium is leached in two counter-current thickening stages using 0.04 molar EDTA. Underflow from the second thickener is fed to 2 trains of 5 CCD thickeners where it is washed with water for EDTA recovery.

adium is precipitated from the solution overflowing the first EDTA thickeners by adjusting the pH with sulphuric acid and adding barium chloride. ubsequently the substantially radium-free solution is treated with lime to remove iron as the hydroxide and to convert all EDTA to the calcium form. fter clarification, the solution is diluted by a factor of three and acidified to precipitate EDTA for recycle. Approximately one third of the inal barren solution is used as wash solution on the uranium CCD thickeners, nd the remainder is diverted to effluent treatment facilities. The barium-adium precipitate filtered from the process solution is fed as a slurry to rums for special waste disposal.

ffluents from the sulphuric acid-EDTA circuit include:

- uranium circuit raffinate
- ore residue slurry, well washed but containing some residual EDTA and radium
- barren solution from the EDTA circuit containing some EDTA and radium

These streams are combined and neutralized to pH 8 with lime prior to disposal n the tailing pond. Pond overflow is treated with barium chloride for radium removal prior to release.

A block flowsheet is shown on Figure 4.2.7.

ii) Selected Design Criteria

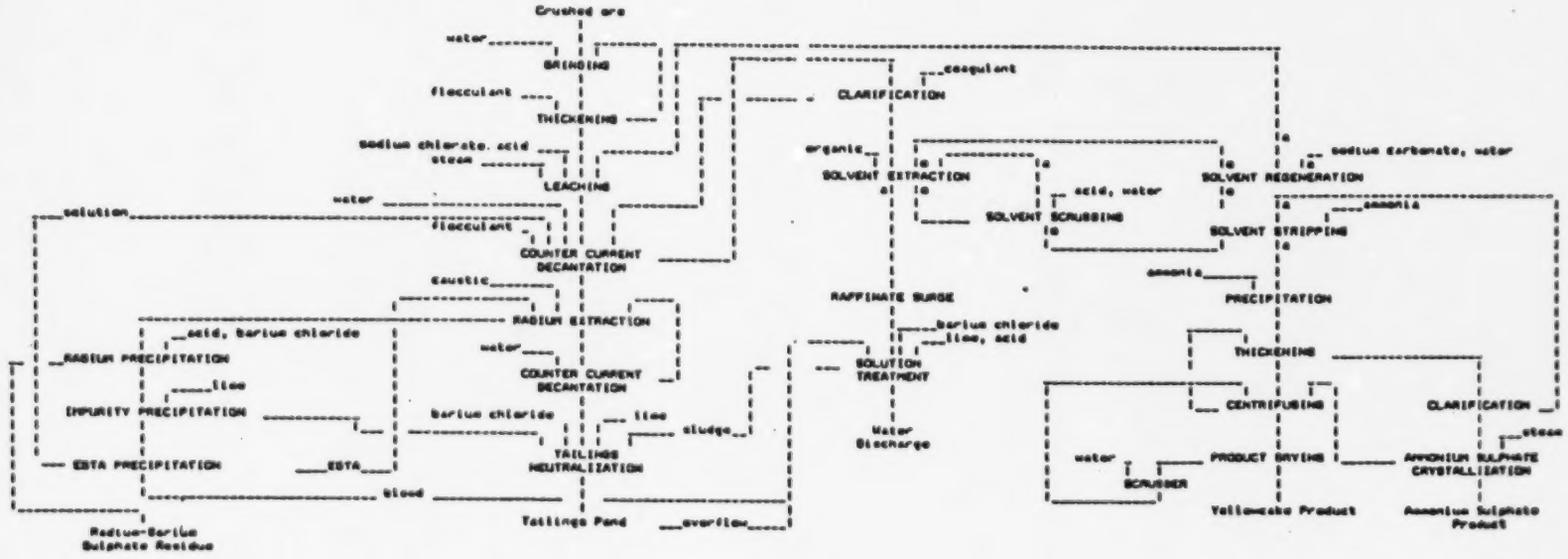
Uranium Circuit (as in Base Case)

EDTA Leaching

EDTA concentration	- 0.04 Molar
Liquid - solid ratio	- 2:1
Radium content of residue	- 2.22 Bq/g
Number of contact stages	- 2
Contact time per stage	- 20 min

Figure 4.2.7

ALTERNATIVE 6 - EDTA LEACH



A.H. ROSE & ASSOCIATES

EDTA Recovery

Number of CCD stages	- 5
Wash ratio	- 2 m^3/t
Soluble loss of EDTA	- 1%

Radium Precipitation

Barium chloride	- 100 mg/L
Radium in barren solution	- 3.7 Bq/L

Impurity Precipitation

Lime consumption	- 5.5 kg/t

EDTA Precipitation

EDTA in barren	- 0.2 g/L
Dilution ratio	- 3:1
pH	- 1.8

Tailings Treatment

- refer to Base Case -

iii) Capital Costs

Summary of Capital Costs	$
EDTA circuit	32,163,000
Tailings treatment	4,576,000
Buildings	12,234,000
Total direct cost	48,973,000
Indirect cost	14,692,000
Total before contingency	63,665,000
Contingency @20%	12,733,000
Total Cost	$76,398,000

Cost for base case	138,100,000
Total capital cost	214,478,000
Amortized cost (20 yrs at 12%)	$27,820,000/a
$/t ore	$10.46
$/kg U_3O_8	$12.31

iii) Operating Costs

Incremental Costs	$
Supervision	52,000
Operating labour	382,000
Maintenance	2,674,000
Process chemicals	47,146,000
Operating supplies	362,000
Power	546,000
Analytical and test laboratories	49,000
Sub-total	51,211,000
Contingency @10%	5,121,000
Total incremental cost	56,332,000
Cost for base case	34,559,000
Total Operating cost	90,981,000
$/t ore	$34.70
$/kg U_3O_8	$40.26

4.2.8 Alternative 7 - HCl Leach

i) Process Description and Flowsheet

Hydrochloric acid leaching testwork has been conducted on Type A ore (M,19) and process economics developed (22). The description below generally follows

that presented by Lummus.

The ore is crushed underground to minus 200 mm and is transported to surface storage silos. It is wet ground to 60% -200 mesh to achieve the required extraction efficiency. Pyrite is floated to decrease environmental problems. The flotation tails are thickened, filtered, and directed to hydrochloric acid leaching.

Pyrite flotation concentrate is roasted in a fluid-bed roaster and the sulphur dioxide gas is cooled in a waste heat boiler, cleaned, and passed to a sulphuric acid plant. Acid production amounts to 190,000 t/year, of which 154,500 t are used in the process. It is assumed that the remainder is marketed. No allowance has been made for revenue from acid. Roaster calcine and collected dust are slurried and pumped to the leaching section.

Pyrite flotation tails and roaster calcine are leached together with hydrochloric acid in a two stage leach system. Since sulphate ions released in the first stage would suppress radium dissolution in the second stage, the first stage leach discharge is filtered. The leach slurry is filtered after the second stage to recover dissolved uranium, thorium, radium, and hydrochloric acid. Acid remaining in the washed filtercake from the second stage is neutralized before the solids are discharged to the tailings pond.

Pregnant solution from first stage filtrate is clarified and treated in a solvent extraction circuit. Uranium is extracted into an organic phase comprised of tri-butyl phosphate and kerosene and subsequently stripped from the organic phase with water. Yellowcake, precipitated with the addition of magnesia and hydrogen peroxide, is thickened, washed, centrifuged, dried and packed. It is considered that product specification would not likely be met with the use of magnesia alone.

Thorium is removed from the solution in a second solvent extraction operation. It is stripped from the solvent with hydrochloric acid solution and the strip solution evaporated to remove excess acid. Thorium is precipitated with the

addition of oxalic acid and the precipitate is washed, dried and stored.

Radium is removed from the solvent extraction barren solution (raffinate) in a series of stirred tanks by co-precipitation with barium sulphate, formed on addition of barium chloride. The barium/radium sludge is separated from the solution using precoat pressure filters and is directed to a special disposal area.

The treated raffinate must be recycled to the process since the chloride content is too high for release to tailings. Excess waste is removed by evaporation, and limestone and lime added to the evaporator bottoms to raise the pH and precipitate metallic constituents as hydroxides. The precipitate is removed by centrifuging and pumped to tailings disposal. Sulphuric acid is added to the concentrate forming hydrochloric acid and gypsum. The gypsum is removed for disposal to tailings; hydrochloric acid is distilled and recycled to leach.

A block flowsheet is shown on Figure 4.2.8.

ii) Selected Design Criteria

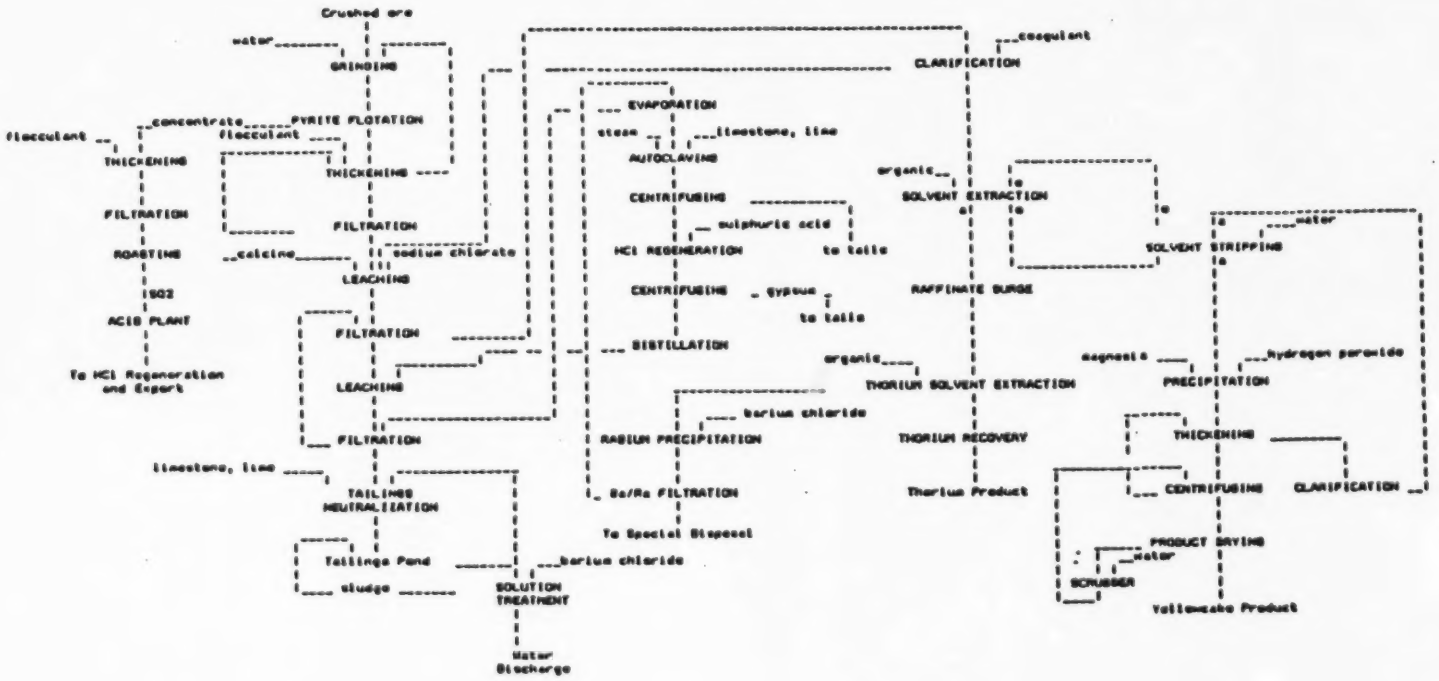
Grind	- 60% passing 200 mesh
Flotation	
Sulphur recovery	- 90%

Leaching

	First Stage	Second Stage
Retention time, hours	6	6
Temperature °C	75	75
Percent solids	50	50
Acidity, kg/t	26	44
Sodium chlorate, kg/t	2.0	2.0
Extraction U_3O_8%	-	98
Th %	-	80

Figure 4.2.8

ALTERNATIVE 7 - HCl LEACH



A.H. ROSE & ASSOCIATES

Ra %	- 95
Uranium Recovery	
Magnesia	- 0.22 kg/kg U_3O_8
Hydrogen peroxide	- 0.18 kg/kg U_3O_8
Radium Removal	
Barium chloride	- 0.1 kg/m^3
Acid Regeneration	
Sulphuric acid	- 75 kg/t

iii) Capital Costs

Capital costs summarized in the following table are based primarily on those previously reported by Lummus (22). Costs for radium recovery and tailings treatment have been estimated independently.

Summary of Capital Costs	$
Grinding, flotation, dewatering	32,070,000
Roasting and acid plant	37,990,000
Leaching, solid/liquid separation	75,030,000
Uranium recovery	12,670,000
Thorium recovery	12,920,000
Radium recovery	2,150,000
Hydrochloric acid regeneration	30,270,000
Tailings neutralization	3,630,000
Utilities	21,650,000
Buildings	38,700,000
Total cost including contingency	$267,080,000
Amortized cost (20 yrs at 12%)	$34,640,000/a
$/t ore	$13.02
$/kg U_3O_8 (2.33×10^6 kg/a)	$14.87

iv) Operating Costs

Summary of Operating Costs	$
Wages and salaries	15,639,000
Maintenance materials	9,962,000
Power	1,331,000
Steam	18,835,000
Water	565,000
Process chemicals	16,319,000
Supplies and minor chemicals	2,362,000
Sub-total	64,563,000
Contingency at 10%	6,456,000
Total operating cost	$71,019,000
$/t ore	$26.70
$/kg U_3O_8 (2.33×10^6 kg/a)	$30.48

4.3 Process Descriptions and Cost Estimates for Ore B Milling Alternatives

4.3.1 Base Case - Sulphuric Acid Leach

i) Process Description and Flowsheet

Atmospheric sulphuric acid leaching as outlined previously is the conventional treatment route for type B ore. Specific details chosen for this development are as follows:

Stockpile ore from an open pit is fed by front-end loader to a jaw crusher, thence to a semi-autogenous mill/ball mill circuit for grinding. Cyclone classified pulp passes directly to leaching without thickening.

The ground ore is leached with sulphuric acid and sodium chlorate in mechanically agitated vessels heated with live steam. Sulphuric acid is

manufactured from purchased sulphur.

The leach discharge solids and solution are separated in a six-stage conventional CCD thickening circuit.

The pregnant liquor is treated in a hopper clarifier and sand filters and fed to a conventional solvent extraction system. Uranium is stripped from the solvent and precipitated with ammonia. Yellowcake is washed in a thickener, centrifuged, and dried in an oil-fired multi-hearth furnace.

Ammonia leaves the system as an ammonium sulphate solution which is directed to a crystallizer for recovery as a fertilizer. Ammonium sulphate is recovered by crystallization from a bleed stream of barren strip solution.

Solvent extraction raffinate is recycled to the CCD circuit as wash solution.

The washed pulp from CCD is neutralized with lime and discharged to the tailings pond. Pond overflow is treated with barium chloride to precipitate barium/radium sulphate which is retained in a settling pond.

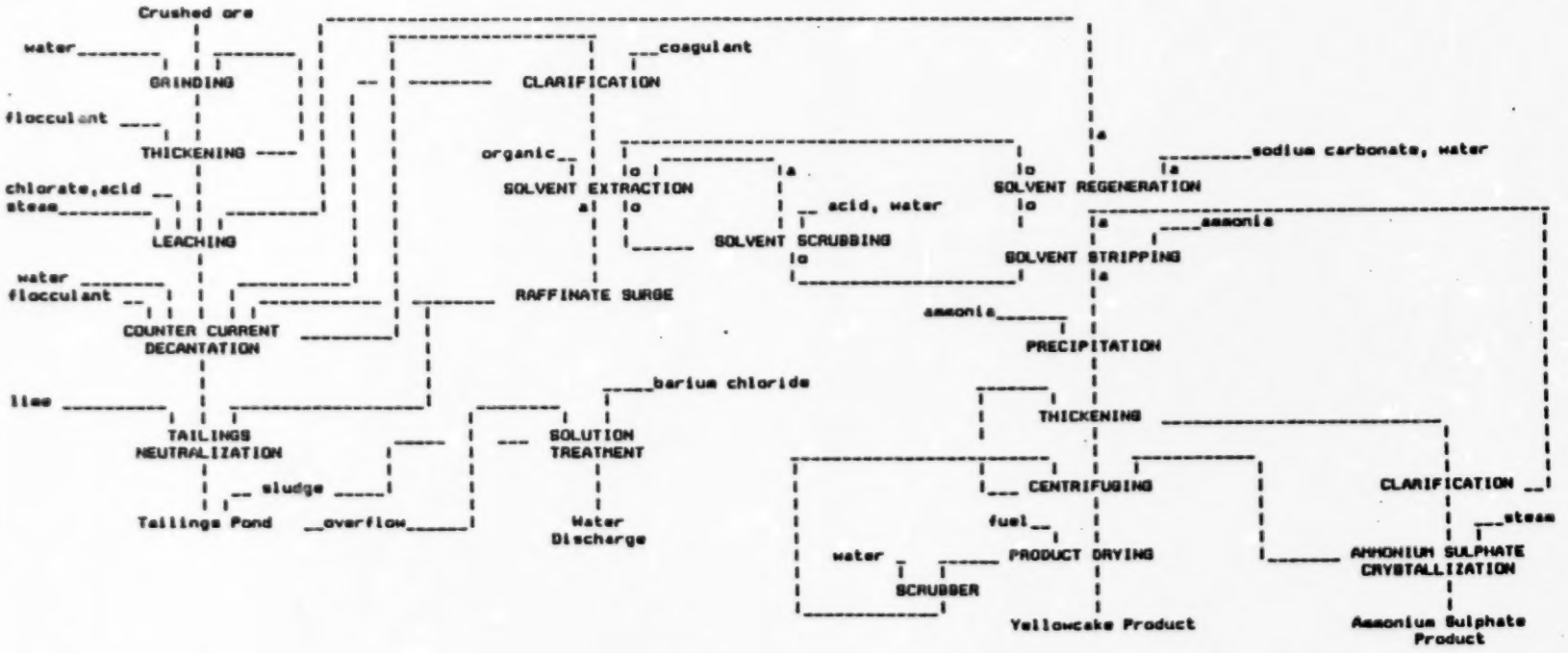
A block flowsheet is shown on Figure 4.3.1.

ii) Selected Design Criteria

Crushing	- -200 mm
Grinding	- 80% passing 48 mesh
Leaching	
Retention time	- 8 hours
Temperature	- 45 °C
Density	- 55% solids
Sulphuric acid	- 65 kg/t
Sodium chlorate	- 3.0 kg/t
Uranium extraction	- 96.5%
Liquid-Solid separation	

Figure 4.3.1

BASECASE - SULPHURIC ACID LEACH



A.H. ROSS & ASSOCIATES

CCD stages	- 6
Soluble loss	- 0.5%
Solvent Extraction	
Pregnant solution	- 3 m^3/t ore
Amine strength	- 3.5%
Stages - extraction	- 4
- scrub	- 1
- strip	- 4
- regeneration	- 1
Ammonia	- 0.4 kg/kg U_3O_8
Uranium Recovery	
Ammonia	- 0.2 kg/kg U_3O_8
Ammonium sulphate concentration	- 150 g/L
Tailings Treatment	
Lime	- 28 kg/t

iii) Capital Costs

Summary of Capital Costs	$
Receiving and crushing	1,880,000
Grinding	5,410,000
Leaching	1,510,000
CCD	11,580,000
Clarification	1,410,000
Solvent extraction	3,770,000
Uranium recovery	3,790,000
Ammonium sulphate recovery	2,020,000
Tailings neutralization	1,190,000
Reagent handling	2,180,000
Utilities	11,470,000
Buildings	13,110,000
Acid Plant	6,400,000

Total direct cost	65,720,000
Indirect cost	26,640,000
Mobile equipment	400,000
Total before contingency	92,760,000
Contingency @ 20%	18,550,000
Total estimated capital cost	$110,090,000
Amortized Cost (20 yrs at 12%)	14,440,000/a
$/t ore	$24.31
$/kg U_3O_8	$6.39

iv) Operating Costs

Summary of Operating Costs	$
Supervision	627,000
Operating labour	2,373,000
Maintenance	3,949,000
Process chemicals	9,195,000
Steam	1,169,000
Power	2,995,000
Supplies	776,000
Analytical and test laboratories	651,000
Sub-Total	21,735,000
Contingency @ 10%	2,174,000
Total annual operating cost	23,909,000
$/t ore	$40.24
$/kg U_3O_8	$10.58

4.3.2 Alternative 1 - HCl Leach

i) Process Description and Flowsheet

Leaching of a type B uranium ore (Rabbit Lake) with hydrochloric acid was studied by CANMET (18). Test work was limited to bench scale, and no detailed flowsheet has appeared. The economics of the process as applied to a type A ore are evaluated in an earlier report by C.E. Lummus, Minerals Division (22). The following description of the process as applied to type B ore is based on the Lummus report.

Ore from an open pit is stockpiled and fed into a primary jaw crusher en route to a primary semi-autogenous/secondary ball mill circuit. Classified pulp is thickened, filtered, and leached with hydrochloric acid in a two-stage leach system. The leach slurry is filtered after each stage. Acid remaining in the filtered and washed solids from the second stage is neutralized before discharging the solids to the tailings pond.

Pregnant solution, abstracted from the first stage filtrate, is clarified and treated in a solvent extraction circuit. Uranium is extracted into an organic phase comprised of tri-butyl phosphate and kerosene. It is stripped from the organic phase with water and precipitated with magnesia and hydrogen peroxide. The slurry is thickened and the solids washed, centrifuged, dried, and packed.

As described in (22), thorium is removed from uranium raffinate in a solvent extraction circuit employing an amine extractant. It is precipitated from hydrochloric acid strip solution with oxalic acid. Barren solution from solvent extraction (raffinate) is directed to the radium removal circuit.

On the addition of barium chloride, radium is substantially removed from the raffinate by a co-precipitation with barium sulphate in a series of stirred tanks. The barium/radium sulphate sludge is removed from the solution with a

precoated pressure filter. It is combined with the thorium precipitate for special disposal.

The chloride content of the treated raffinate is too high for release to the environment. Accordingly, the metallic constituents are removed from the liquor and the chloride converted to hydrochloric acid in a regenerating circuit. Excess water is first removed by evaporation; the low acid distillate is utilized as filter wash in the leaching circuit. Evaporator bottoms are treated with limestone and lime to precipitate metal hydroxides. The precipitate is removed by centrifuging. Sulphuric acid is added to the concentrate to produce hydrochloric acid and gypsum. The gypsum and hydroxides are sent to tailings. The hydrochloric acid is distilled and recycled to leaching.

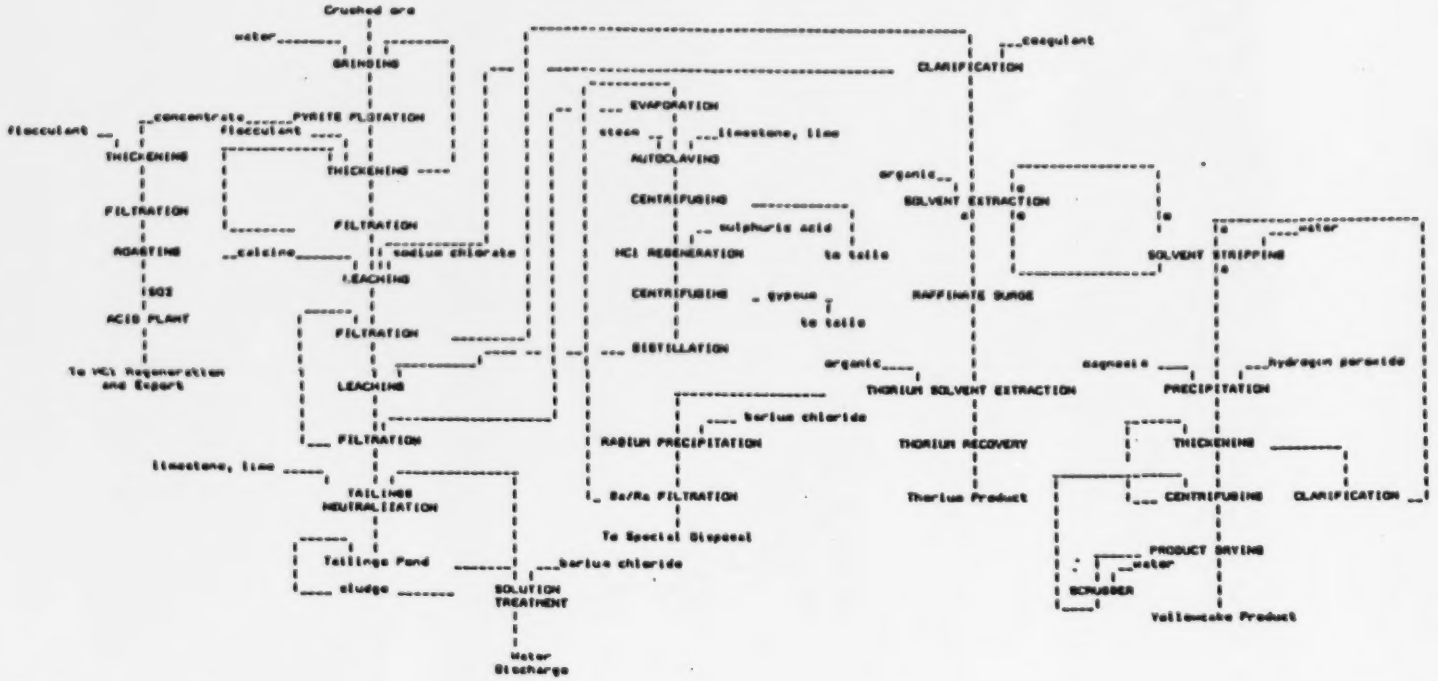
A block flowsheet is provided in Figure 4.3.2.

ii) Selected Design Criteria

Grinding	- 50% passing 200 mesh	
Leaching		
	First Stage	Second Stage
Retention time, hours	10	6
Temperature °C	60	60
Percent solids	50	50
Hydrochloric acid, kg/t	88	66
Sodium chlorate, kg/t	2	-
Percent extraction: U_3O_8	96	98.5
Ra	77	88
Th	-	80
U_3O_8 Precipitation		
Magnesia	- 0.22 kg/kg U_3O_8	
Hydrogen peroxide	- 0.18 kg/kg U_3O_8	
Radium removal		
Barium chloride	- 0.1 kg/m^3	

Figure 4.3.2

ALTERNATIVE 1 - HCl LEACH



A.H. ROSS & ASSOCIATES

Acid regeneration

Sulphuric acid - 120 kg/t

iii) Capital Costs

Capital costs are summarized in the following table. The cost of leaching, uranium recovery, thorium removal, hydrochloric acid regeneration and much of the utilities and buildings are based on the Lummus report. Costs of the other areas were estimated independently.

Summary of Capital Costs	$
Receiving and crushing	2,930,000
Grinding	12,970,000
Leaching	34,030,000
Uranium recovery	11,210,000
Thorium recovery	3,350,000
Radium removal	1,850,000
Hydrochloric acid regeneration	20,270,000
Tailings neutralization	1,890,000
Utilities	18,400,000
Acid Plant	8,390,000
Buildings	21,870,000
Mobile equipment	400,000
Sub-total	137,560,000
Contingency @ 20%	27,512,000
Total	165,072,000
Amortized Cost (20 yrs at 12 %)	$21,410,000/a
$/t ore	$36.04
$ kg U_3O_8 (2.31×10^6)	$9.27

iv) Operating Costs

Summary of Operating Costs	$
Wages and salaries	9,482,000
Maintenance materials	5,777,000
Power	1,021,000
Steam	4,100,000
Water	126,000
Process chemicals	13,351,000
Supplies	1,048,000
Sub-total	34,905,000
Contingency @ 10%	3,490,000
Total Operating cost	38,396,000
$/t ore	$64.64
$/kg U_3O_8 (2.31×10^6 kg/a)	$16.62

4.3.3 Alternative 2 - HCl + Cl_2 Leach

i) Process Description and Flowsheet

Laboratory results have been reported on the attack of dry chlorine gas on type A ore but not on Type B. The economics of the process as applied to type C have been studied and the following process description applicable to Ore B has been adapted from that study.

Open pit ore from a stockpile passes through a primary jaw crusher and a primary semi-autogenous mill/secondary ball mill circuit. The classified pulp is thickened, filtered, and conveyed to a fluid bed dryer. To obtain high chlorination efficiency the ore must be bone dry.

The dried product is chlorinated in two fluid bed reactor in series followed by a cooler. Fluidization is achieved with a mixture of nitrogen and chlorine. The exit gases contain chlorides of uranium, thorium, and iron.

Radium is not volatilized. The gases are passed to a chlorine regenerating reactor in which oxygen reacts with the volatile metal chlorides and elemental chlorine is released. The metals are converted to oxides which are solids at the temperature employed.

The solids are removed from the gas and are conveyed to a sodium carbonate leaching circuit to recover uranium. The gas is cooled and scrubbed with cold water to recover the chlorine. The scrub water is heated to distill the chlorine, which is dried, liquified by compression and refrigeration, and recycled to the chlorination circuit.

Calcined solids from the chlorination reactor are leached with dilute hydrochloric acid to extract radium and unvolatilized uranium. The solution is recovered by filtration and is clarified before being processed by solvent extraction to recover the uranium. Uranium is stripped from the organic phase with water and precipitated with magnesia and hydrogen peroxide.

The oxides from chlorination are processed through a carbonate leach and the uranium recovered as sodium diuranate. To meet refinery specifications the sodium diuranate is dissolved in acid and precipitated with the strip solution from solvent extraction. The yellowcake slurry is thickened, washed, centrifuged, dried and packed.

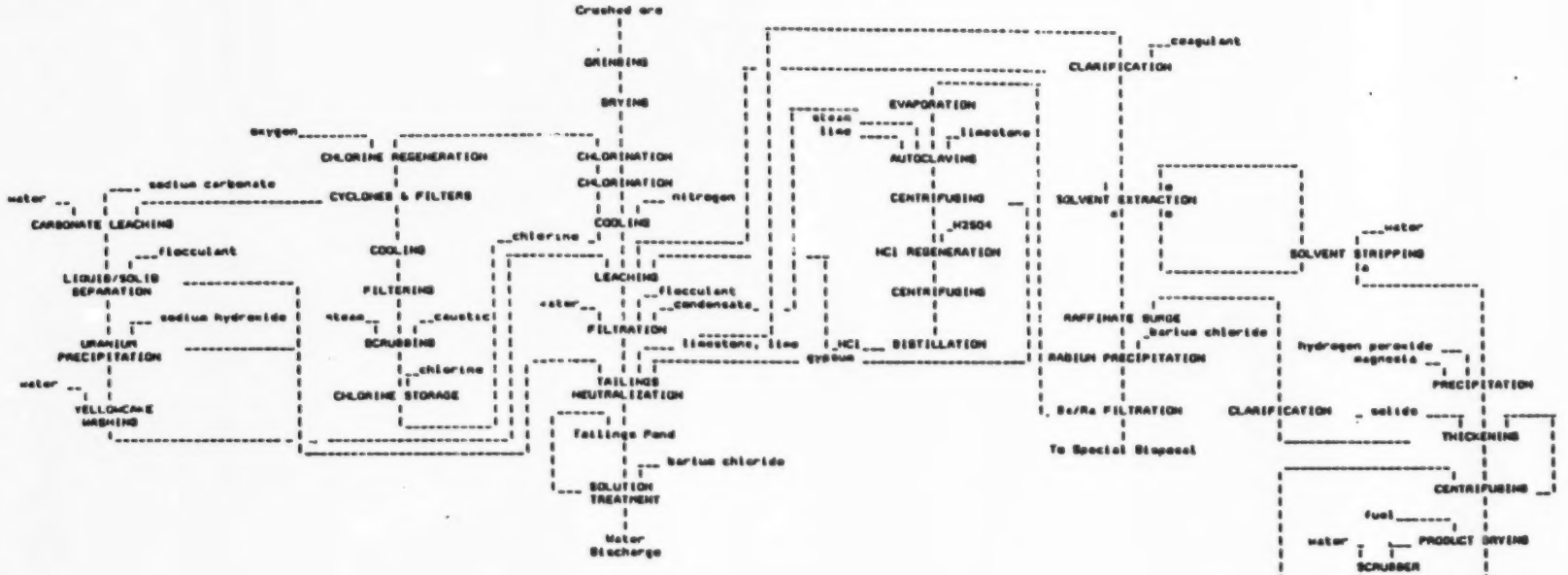
Radium removal from the raffinate and regeneration of the treated solution follows the procedures described in Section 4.3.2.

Thorium is not effectively extracted and passes to tailings with the other ore solids.

A block flownseet is provided in Figure 4.3.3.

Figure 4.3.3

ALTERNATIVE 2 - HCl + Cl_2 LEACH



A.H. ROSS & ASSOCIATES

ii) Selected Design Criteria

Grinding	- 60% passing 200 mesh
Chlorination	
Temperature: first stage	- 400°C
second stage	- 600°C
Volatilization: U_3O_8	- 60%
Th	- 40%
Ra	- 0%
Hydrochloric Acid Leaching	
Extraction: U_3O_8	- 36%
Th	- 5%
Ra	- 94%
Acid concentration	- 0.05 M
Time	- 1 hour
Temperature	- 80°C
Pregnant solution flow	- 1.0 m^3/t
Uranium precipitation	
Magnesia	- 0.22 kg/kg U_3O_8
Hydrogen peroxide	- 0.18 kg/kg U_3O_8
Radium removal	
Barium chloride	- 0.1 kg/m^3
Hydrochloric acid regeneration	
Sulphuric acid	- 2.5 kg/t

iii) Capital Costs

Estimated capital costs for drying, chlorination and chlorine regeneration, leaching, uranium recovery, hydrochloric acid regeneration, and much of utilities and buildings are based on reference (22). Costs for receiving and crushing,grinding, radium removal, and tailings treatment were estimated independently.

Summary of Capital Cost	$
Receiving and Crushing	2,930,000
Grinding	12,970,000
Dryer	5,400,000
Chlorination and Chlorine Regeneration	57,000,000
Leaching	18,690,000
Uranium recovery	11,210,000
Radium removal	1,890,000
Hydrochloric acid regeneration	12,000,000
Tailings neutralization	1,890,000
Utilities	26,690,000
Buildings	41,270,000
Mobile equipment	160,000
Sub-total	$192,100,000
Contingency @ 20%	38,420,000
Total	230,520,000
Amortized cost (20 yrs at 12%)	$29,898,000/a
$/t ore	$50.33
$/kg U_3O_8	$13.23

iv) Operating Costs

Summary of Operating Costs	$
Wages and Salaries	16,675,000
Maintenance materials	7,767,000
Power	5,280,000
Steam	11,000,000
Process chemicals	2,969,000
Water	1,289,000
Operating supplies	1,070,000
Sub-total	46,050,000

Contingency @10%	4,605,000
Total operating cost	$50,655,000
$/t ore	$85.28
$/kg U_3O_8	$22.41

4.3.4 Alternative 3 - Carbonate Leach

i) Process Description and Flowsheet

Ore crushing, grinding and thickening are as described in the base case except that grinding mill sizes are increased to yield a ground product of 70% minus 200 mesh. Grinding is done in recycled carbonate solution.

Leaching is conducted in agitated tanks using air as an oxidant. Because of the 80°C operating temperature, pulp to pulp, and steam to pulp heat exchange equipment is provided. Liquid/solid separation is accomplished with two stages of drum filters. Barren solution is used as a wash on the second stage. Second stage filtrate is recycled to the grinding circuit.

Pregnant solution is clarified in a hopper clarifier and sand filters. Clarified pregnant solution is partially evaporated in a multiple-effect evaporator to minimize soluble loss and maintain a solution balance. Subsequently, lime is added for control of sodium ion concentration and the solution is again clarified.

Uranium is precipitated as sodium diuranate and the barren solution is recycled to primary filtration. The centrifuged yellowcake does not meet refinery specifications and is re-dissolved in sulphuric acid. A final yellowcake product is produced by precipitation of uranyl peroxide using hydrogen peroxide, and drying in a low temperature dryer.

The secondary filtercake is repulped with tailings recycle water, and pumped to the tailings pond.

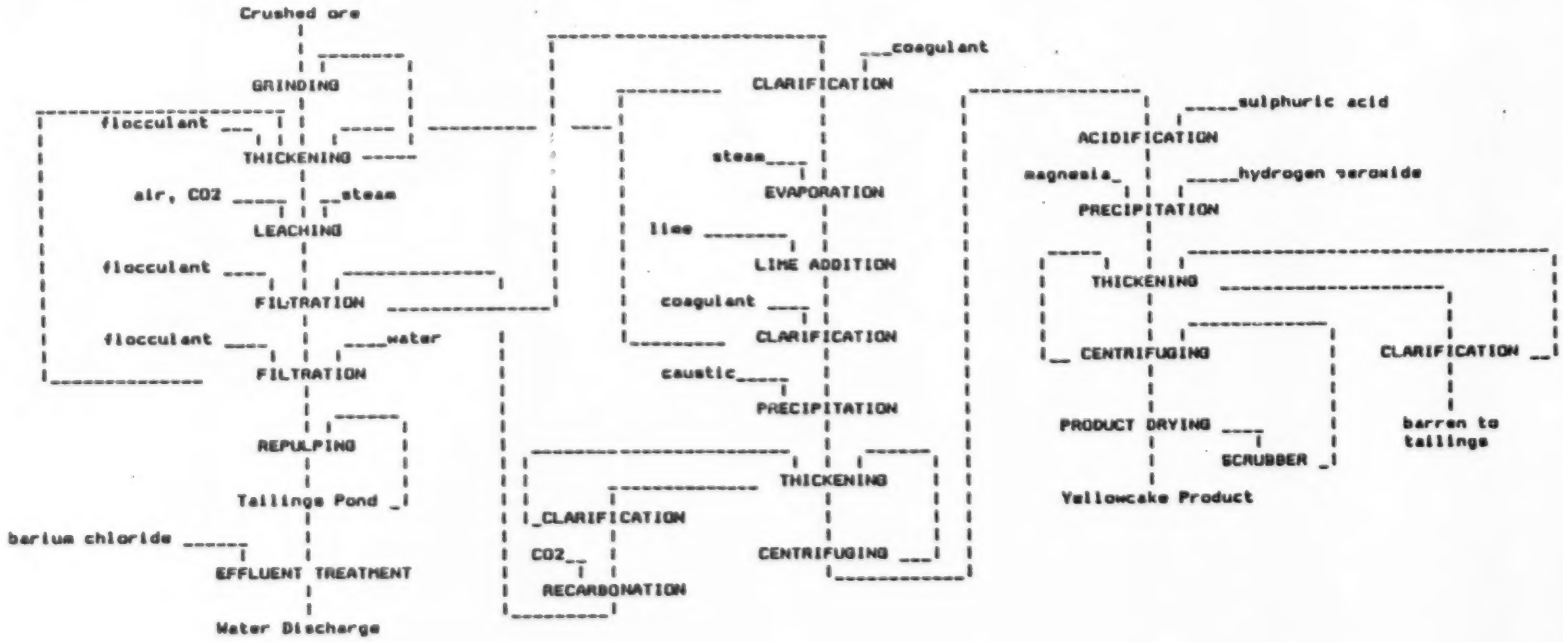
A block flowsheet is provided in Figure 4.3.4.

ii) Selected Design Criteria

Crushing	- minus 200 mm
Grinding	
Product size	- minus 65 mesh
Thickener unit area	- 0.5 m^2.d/t
Leaching	
Retention time	- 48 h
Pulp density	- 50% solids
Temperature	- 80°C
Sodium carbonate concentration	- 50 g/L (total)
Uranium extraction	- 96.5%
Liquid-Solid Separation	
Equipment	- drum filter
Stages	- 2
Flocculant	- 0.4 kg/t
Soluble loss	- 0.5%
Lime Plant	
Retention time	- 30 min.
Uranium Recovery	
Preg-ore ratio	- 0.65:1
Excess Caustic	- 5 g/L
	- 6 kg/t
Acid consumption	- 3 kg/t
Hydrogen peroxide	- 1 kg/t
Magnesia	- 0.75 kg/t

Figure 4.3.4

ALTERNATIVE 3 - CARBONATE LEACH



A.H. ROSS & ASSOCIATES

iii) Capital Costs

Summary of Capital Costs	$
Ore Receiving and Crushing	1,216,000
Grinding and Thickening	7,768,000
Leaching	5,097,000
Filtration	11,580,000
Clarification, Evaporation, Lime Plant	3,278,000
Uranium Recovery	7,284,000
Tailings Pipeline	579,000
Reagent Handling	1,031,000
Utilities	14,160,000
Buildings	12,856,000
Total direct costs	64,849,000
Indirect costs	29,182,000
Mobile Equipment	400,000
Total	94,431,000
Contingency @20%	18,886,000
Total estimated capital cost	$113,317,000
Amortized Cost (20 yrs at 12%)	14,697,000/a
$/t ore	$24.74
$/kg U_3O_8	$6.50

iv) Operating Costs

Summary of Operating Costs	$
Supervision	618,000
Operating Labour	2,772,000
Maintenance	5,666,000
Reagents	5,337,000
Operating Supplies	895,000
Power	3,992,000

Fuel	2,894,000
Analytical and Test Laboratories	624,000
Sub-total	22,798,000
Contingency @10%	2,280,000
Total operating cost	$25,078,000
$/t ore	$42.22
$/kg U_3O_8	$11.01

4.4 Process Descriptions and Cost Estimates for Ore C Milling Alternatives

4.4.1 Base Case - Sulphuric Acid Leach

i) Process Description and Flowsheet

The conventional sulphuric acid leach process has been outlined in Section 4.1. The following is a more detailed description specific to treatment of type ore C.

Open pit ore is drawn from blending stockpiles by front-end loader and crushed in a jaw crusher. Crushed ore is fed directly to a semi-autogenous grinding (SAG) mill via a small surge bin.

The SAG mill is in closed circuit with a cyclone and operated for one 12 hour shift per day to produce a minus 35 mesh product. Ground pulp is stored in surge tanks to allow 24 hour operation of the remainder of the mill.

Uranium is leached in a single stage system using one line of autoclaves and oxygen gas as the oxidant. Uranium extraction is 98.5%. Sulphuric acid is generated in a sulphur burning acid plant and oxygen is provided by a pressure swing adsorption oxygen plant.

Leach discharge pulp is washed in a CCD circuit consisting of one line of 7 conventional thickeners. Pregnant liquor totals 4.75 m^3/t of ore, and soluble losses (all sources) is 0.5%.

Pregnant solution is clarified in a hopper clarifier and sand filters. Uranium is recovered from solution by solvent extraction using a tertiary amine. Conventional mixer-settlers are used. Uranium is stripped from the solvent with ammonium sulphate and is precipitated as ammonium diuranate. Solvent extraction includes a scrub stage primarily for arsenic control. Yellowcake is washed in a thickener and centrifuge, and dried in an oil-fired multi-hearth dryer.

Ammonia is recovered as ammonium sulphate from a bleed stream of barren solution.

Raffinate, along with tailings pond water and other minor liquid effluent streams, is neutralized with lime and treated with barium chloride prior to discharge to the environment. Sludges from this treatment are combined with CCD tailings, neutralized with lime and discharged to the tailings pond.

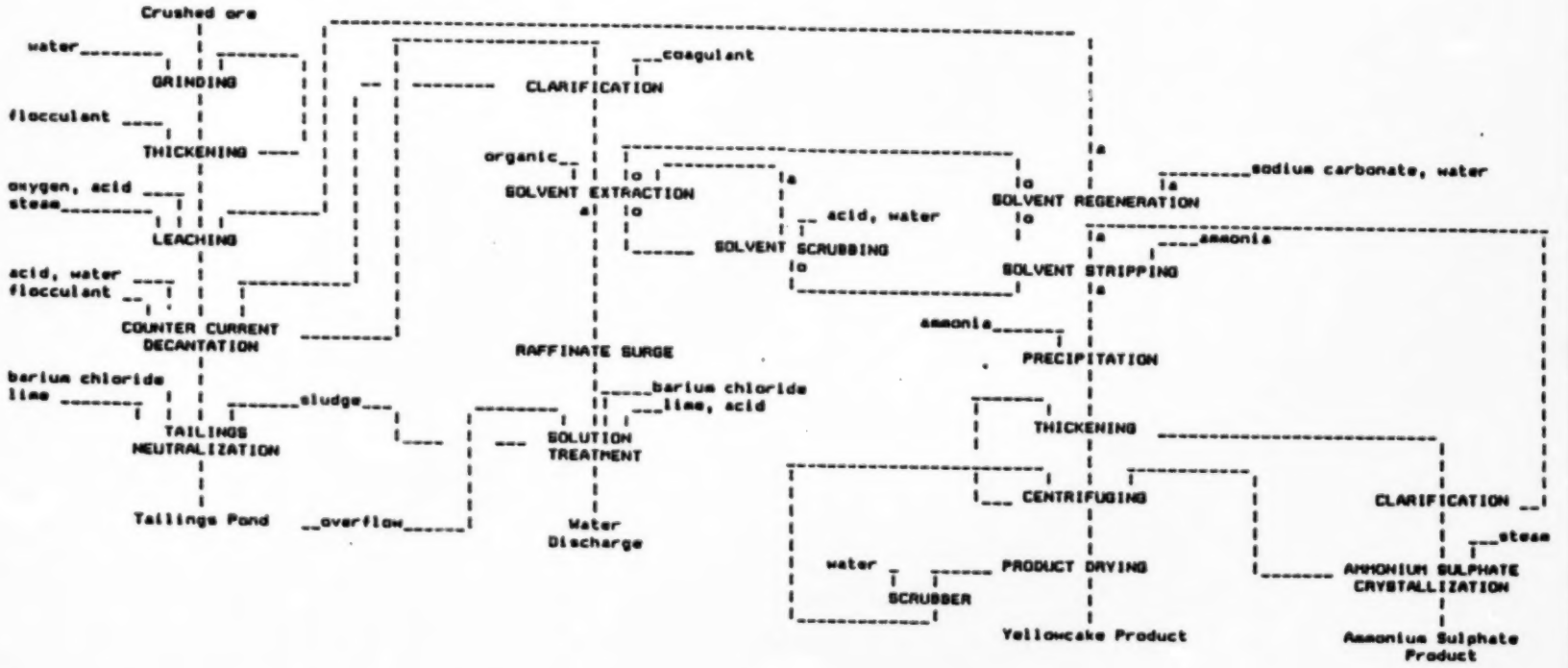
A block flowsheet is shown on Figure 4.4.1.

ii) Selected Design Criteria

Crushing	- minus 200 mm
Grinding	
Product Size	- -35 mesh
Leaching	
Retention time	- 6 h
Equipment	- horizontal autoclaves
Feed pulp density	- 50% solids
Ore weight loss	- 15%
Pressure	- 650 kPa
Temperature	- 70°C
Acid consumption	- 100 kg/t
Uranium extraction	- 98.5%
Liquid-Solids Separation	
Number of CCD stages	- 7

FIGURE 4.4.1

BASE CASE - SULPHURIC ACID LEACH



A.H. ROSE & ASSOCIATES

Soluble loss	- 0.5%
Solvent extraction	
Pregnant solution	- 4.75 m^3/t
Stages - extraction	- 4
stripping	- 4
scrub	- 2
regeneration	- 1
Amine strength	- 5%
Ammonia	- 0.4 kg/kg U_3O_8
Uranium Recovery	
Ammonium sulphate concentration	- 150 g/L
Ammonia	- 0.2 kg/kg U_3O_8
Tailings Treatment	
Lime	- 130 kg/t
Barium chloride	- 0.6 kg/t

iii) Capital Costs

Summary of Capital Costs	$
Ore receiving and crushing	1,477,000
Grinding	2,009,000
Leaching	4,248,000
CCD Thickening	3,747,000
Clarification	786,000
Solvent Extraction	2,140,000
Uranium Recovery	4,065,000
Tailings Treatment	9,653,000
Ammonium Sulphate Recovery	2,484,000
Reagent Handling	1,298,000
Acid Plant c/w building	5,516,000
Oxygen Plant c/w building	3,234,000

Utilities	9,384,000
Process Buildings	11,105,000
Total Direct Cost	61,146,000
Indirect costs	21,838,000
Mobile Equipment	160,000
Total before contingency	83,144,000
Contingency @20%	16,629,000
Total estimated capital cost	99,773,000
Amortized cost (20 yrs at 12%)	$12,941,000/a
$/t ore	$112.53
$/kg U_3O_8	$5.73

iv) Operating Costs

Summary of Operating Costs	$
Supervision	735,000
Operating Labour	2,454,000
Maintenance	3,492,000
Reagents	2,254,000
Operating Supplies	227,000
Power	3,276,000
Fuel	535,000
Analytical and Test Laboratories	713,000
Sub-total	13,686,000
Contingency @10%	1,369,000
Total operating cost	$15,055,000
$/t ore	$130.91
$/kg U_3O_8	$6.66

4.4.2 Alternative 1 - HCl Leach

i) Process Description and Flowsheet

No test reports on hydrochloric acid leaching of type C ore are available. The description below is based on inference from hydrochloric leach results reported for ores A and B, and chlorine assisted leaching testwork on Key Lake ore.

Except for the much reduced throughput of ore, the process closely follows that of 4.3.2. The possible effect of various sulphide and arsenide minerals in the ore on radium dissolution is not known. The presence in solution of significant amounts of nickel and arsenic, which add to the complexity of effluent disposal in the sulphuric acid flowsheet, are considered to have little effect with the hydrochloric flowsheet as they would be removed with other metallics in the acid regeneration step.

A block flowsheet is provided in Figure 4.4.2.

ii) Selected Design Criteria

Grinding - 55% passing 200 mesh

Leaching

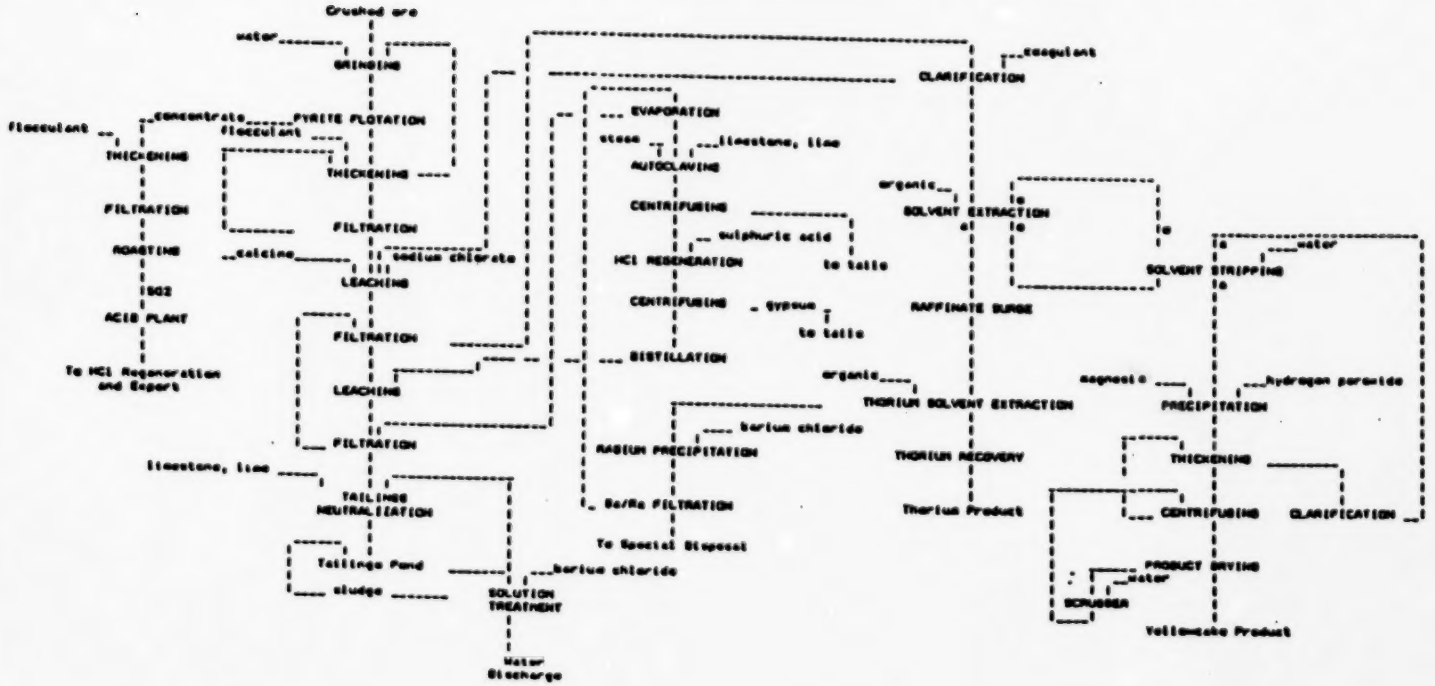
	First Stage	Second Stage
Retention time, hours	10	6
Temperature °C	80	65
Percent solids	50	50
Hydrochloric acid, kg/t	88	44
Sodium chlorate	11.5	2
Percent extraction; U_3O_8	96	99
Ra 226	82	96
Th	-	80

Uranium precipitation

Magnesia - 0.22 kg/kg U_3O_8

Figure 4.4.2

ALTERNATIVE 1 - HCl LEACH



A.H. ROSE & ASSOCIATES

Hydrogen peroxide	- 0.18 kg/kg U_3O_8
Radium removal	
Barium chloride	- 0.1 kg/m^3
Acid regeneration	
Sulphuric acid	- 160 kg/t

ii) Capital Costs

Capital costs are summarized in the following table. The estimated cost of grinding, leaching, uranium recovery, hydrochloric acid generation, and much of utilities and building are based on the costs reported in reference (22). Cost of other areas were estimated independently.

Summary of Capital Costs	$
Receiving and crushing	2,140,000
Grinding	7,730,000
Leaching	16,000,000
Uranium recovery	10,340,000
Radium recovery	1,440,000
Thorium removal	2,000,000
Hydrochloric acid regeneration	13,000,000
Tailings Neutralization	880,000
Utilities	14,860,000
Acid Plant	4,800,000
Buildings	22,010,000
Mobile equipment	160,000
Sub-total	95,360,000
Contingency @ 20%	19,070,000
Total	$114,430,000
Amortized cost (20 yrs at 12%)	$14,842,000/a
$/t ore	$129.06
$/kg U_3O_8 (2.27×10^6 kg/a)	$6.54

iv) Operating Costs

Summary of Operating Costs	$
Wages and salaries	7,983,000
Maintenance materials	3,290,000
Power	2,530,000
Steam	6,843,000
Water	50,000
Process chemicals	5,011,000
Operating supplies	600,000
Sub-total	26,307,000
Contingency @ 10%	2,631,000
Total	$28,938,000
$/t ore	$251.63
$/kg U_3O_8 (2.27×10^6 kg/a)	$12.75

4.4.3 Alternative 2 - HCl + Cl_2 Leach

i) Process Description and Flowsheet

The economics of the dry chlorination process as applied to type C ore have been studied (22) but no test results have been reported. The description below is based on the economic study.

Open pit ore from a stockpile passes through a primary jaw crusher and primary semi-autogenous mill/secondary ball mill circuit. The classified pulp is thickened, filtered, and conveyed to a fluid bed dryer. To obtain high chlorination efficiency the ore must be bone dry.

The dried product is chlorinated in two fluid bed reactors in series followed by a cooler. Fluidization is achieved with a mixture of nitrogen and

chlorine. The exit gases contain chlorides of uranium, thorium, iron, nickel, and arsenic; radium is not volatilized. The gases are passed to a chlorine regenerating reactor in which oxygen reacts with the volatile metal chlorides and elemental chlorine is released. The metals are converted to oxides, which, with the exception of arsenic oxide, are solids at the temperatures employed.

The solid oxides, containing over half the uranium in the feed ore, are removed from the gas and are conveyed to a sodium carbonate leaching circuit to recover uranium. The gas is subsequently cooled and the arsenic oxide condensed, collected, and packaged for sale or disposal.

The cleaned off-gas is treated to recover chlorine by scrubbing in cold water. Traces of chlorine remaining in the scrubbed off-gas are removed by caustic scrubbing. The chlorine scrub water is heated to distill the chlorine, which is dried, liquified by compression and refrigeration, and recycled to the chlorination circuit.

Uranium is removed from the metal oxides by leaching with sodium carbonate. It is precipitated from the leach liquor with caustic soda. In order to remove sodium and meet refinery specifications, the solids, after washing, are added to the hydrochloric acid leach system. The residue from carbonate leach is directed to tailings.

Calcined solids from the chlorination reactors are leached with hydrochloric acid to extract radium and unvolatilized uranium. In conformation to testwork conditions for ore A (25), but in distinction to the economic study (22), a single-stage leach system employing 0.05 M HCl is used. Uranium is extracted from hydrochloric acid leach solution by solvent extraction. The sodium diuranate precipitate from the carbonate leach system is dissolved in the strip solution and uranium from both circuits precipitated with magnesia and hydrogen peroxide. The yellowcake is thickened, washed, centrifuged, dried and packed.

Radium is removed from the raffinate by co-precipitation with barium sulphate on the addition of barium chloride. The barium/radium sludge is separated from the solution for special disposal.

The treated solution is passed to hydrochloric acid regeneration as described in 4.3.2 for removal of metallics and preparation of hydrochloric acid for reuse. The metal hydroxides are passed to tailings.

Volatized thorium reports to carbonate leach residue, and unvolatilized thorium splits between ore tailings and the metal hydroxides from hydrochloric acid regeneration. A method for collection of the thorium for special disposal is not included.

The residue from hydrochloric acid leaching is neutralized after filtration and pumped to the tailings pond.

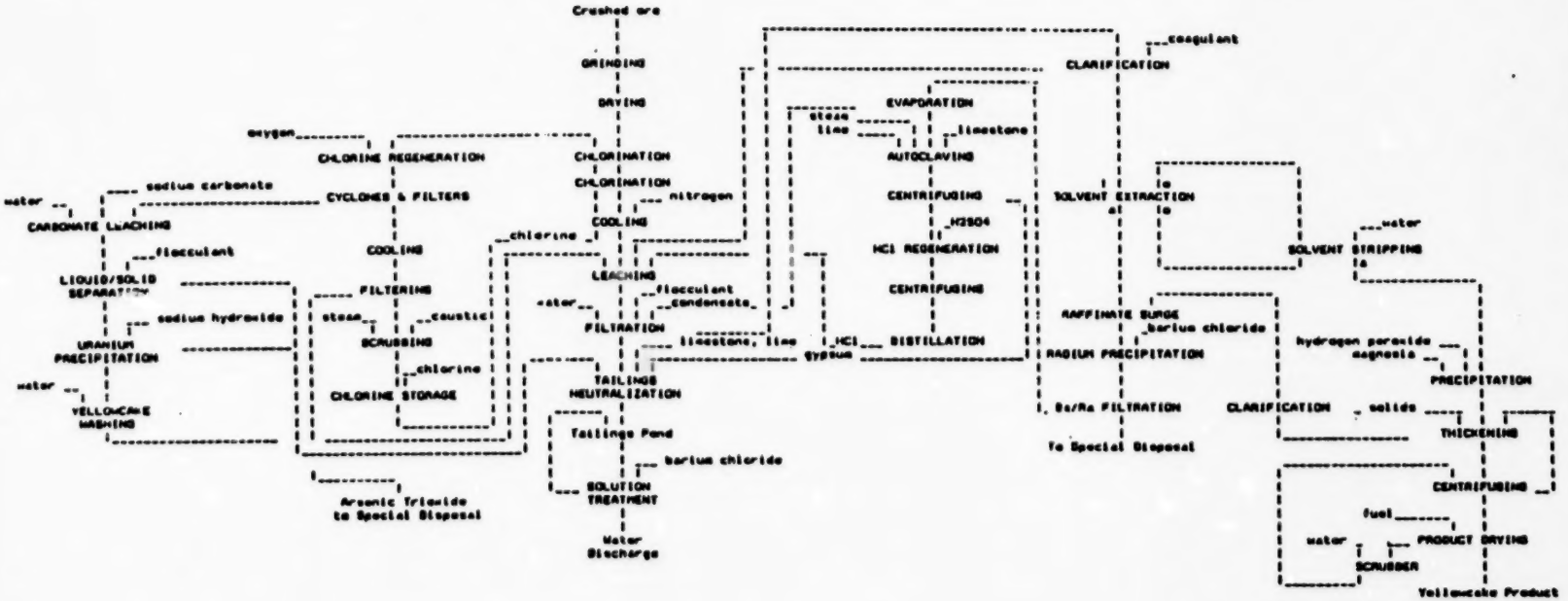
A block flowsheet is provided in Figure 4.4.3.

ii) Selected Design Criteria

Grinding		- 60% passing 200 mesh
Chlorination		
Temperature:	first stage	- 400°C
	second stage	- 600°C
Volatilization:	U_3O_8	- 60%
	Th	- 40%
	Ra	- 0%
Hydrochloric Acid Leaching		
Temperature		- 80°C
Extraction:	U_3O_8	- 36%
	Th	- 5%
	Ra	- 94%
Acid concentration		- 0.05 M
Time		- 1 hour

Figure 4.4.3

ALTERNATIVE 2 - $HCl + Cl_2$ LEACH



A.H. ROSS & ASSOCIATES

Uranium precipitation	
Magnesia	- 0.22 kg/kg U_3O_8
Hydrogen peroxide	- 0.18 kg/kg U_3O_8
Radium removal	
Barium chloride	-0.1 kg/m^3
Hydrochloric acid regeneration	
Sulphuric acid	- 2.5 kg/t

iii) Capital Costs

Estimated capital costs for grinding, drying, chlorination and chlorine regeneration, leaching, uranium recovery, hydrochloric acid regeneration, and much of utilities and buildings are based on costs reported in reference (22). Receiving and crushing, radium removal, and tailings treatment were estimated independently.

Summary of Capital Costs	$
Receiving and crushing	2,140,000
Grinding	7,730,000
Drying	2,310,000
Chlorination and chlorine regeneration	25,520,000
Leaching	11,540,000
Uranium recovery	10,340,00
Radium removal	1,440,000
Hydrochloric acid regeneration	6,000,000
Tailings neutralization	880,000
Utilities	19,060,000
Buildings	24,110,000
Mobile equipment	160,000
Sub-total	$111,230,000
Contingency @ 20%	22,250,000
Total	133,480,000

Amortized Cost (20 yrs at 12%)	$17,312,000/a
$/t ore	$150.54
$/kg U_3O_8	$7.87

v) Operating Costs

Summary of Operating Costs	$
Wages and salaries	13,476,000
Maintenance materials	4,270,000
Power	2,805,000
Steam	7,560,000
Water	250,000
Process chemicals	2,487,000
Operating supplies	560,000
Sub-total	31,408,000
Contingency @ 10%	3,141,000
Total operating costs	34,549,000
$/t ore	$300.43
$/kg U_3O_8	$15.70

4.5 Discussion and Summary

4.5.1 General

Table 4.5.1 provides a summary of the total capital cost, capital cost amortized at 12 % over 20 years, capital cost per kg U_3O_8 total operating cost, operating cost per ton of ore, operating cost/kg U_3O_8 and the total cost per kg U_3O_8 for each of the ore-type milling alternatives.

For all three ore types, the base case milling cost is the least expensive of the various milling alternatives. This is not overly surprising as this is for the most part the process selected by most operators. Process options

Table 4.5.1

CAPITAL AND OPERATING COST SUMMARY FOR ORE MILLING ALTERNATIVES

Ore Type	Milling Alternative	Annual Production (10^6 kg U_3O_8)	Capital Costs			Operating Costs			Total Cost $/kg U_3O_8	Per cent Difference
			Total ($10^6)	Amortized ($10^6/a)	$/kg U_3O_8	Total ($10^6/a)	$/t ore	$/kg U_3O_8		
A	Base Case	2.26	138.1	17.91	7.93	34.56	12.99	15.29	23.22	-
A	Alt. 1 - Pyrite Flotation	2.26	147.3	19.11	8.46	37.79	14.21	16.72	25.18	+8
A	Alt. 2 - Acid Production	2.26	188.5	24.45	10.82	36.00	13.53	15.93	26.75	+15
A	Alt. 3 - Bulk Flotation	2.26	152.5	19.78	8.75	43.22	16.25	19.12	27.87	+20
A	Alt. 4 - Autogenous Leach	2.26	153.2	19.87	8.79	38.36	14.42	16.97	25.76	+11
A	Alt. 5 - Autogenous Leach	2.26	180.9	23.47	10.38	36.40	13.68	16.10	26.93	+14
A	Alt. 6 - EDTA Leach	2.26	214.5	27.82	12.31	90.98	34.20	40.26	52.57	+126
A	Alt. 7 - HCl Leach	2.33	267.1	34.64	14.87	71.0	26.70	30.48	45.35	+95
B	Base Case	2.26	111.3	14.44	6.39	23.91	40.25	10.58	16.90	-
B	Alt. 1 HCl Leach	2.31	165.1	21.41	9.27	38.40	64.64	16.62	25.89	+53
B	Alt. 2 HCl + Cl_2 Leach	2.26	230.5	29.90	13.23	50.66	85.28	22.41	35.64	+111
B	Alt. 3 Carbonate Leach	2.26	113.3	14.70	6.50	25.08	42.22	11.10	17.60	+4
C	Base Case	2.26	99.77	12.94	5.73	15.06	130.91	6.66	12.39	-
C	Alt. 1 HCl Leach	2.27	114.4	14.84	6.54	28.94	251.63	12.75	19.29	+56
C	Alt. 2 - HCl + Cl_2 Leach	2.20	133.5	17.31	7.87	34.55	300	15.70	23.57	+90

including EDTA leach, HCl leach or HCl + Cl_2 leach are very much more expensive than the base cases, with increased costs of 53 to 126 per cent. All other alternatives are within 20 per cent of the base case costs.

Using the base case to compare the milling cost for the three ore grades, operating cost per tonne of ore increases with ore grade and overall uranium production cost decreases with ore grade.

The general relationships are as follows:

operating cost ($/tonne ore) $\propto$ (ore grade) $^{0.75}$

total uranium milling cost ($/kg U_3O_8) $\propto$ (1/ore grade)$^{0.2}$

A brief review of the major cost items for each alternative as compared with the base case is provided in the following section.

Operating costs for the Base Case facility are 2 times the amortized capital cost for milling type A ore. The primary factors in the operating cost are equipment maintenance, process chemicals and fuel. Alternatives such as pyrite removal add marginally to capital costs ($0.50 kg U_3O_8) but contribute to higher operating costs ($1.50/kg U_3O_8). Therefore if pyrite removal could be deferred as a management option, its impact on product cost would be reduced. Given that removal of pyrite is only justified from the surface zones this alternative may have some positive benefits. Environmental benefits would likely need to be substantial to justify the increased capital costs associated with the other processing alternatives.

As with Ore A, operating costs are substantially higher than the amortized capital costs for the base case. Both capital and operating costs are lower for type B ore. Process chemicals and maintenance are the major operating costs. Carbonate leaching is a competitive milling process with estimated total costs only marginally higher than the base case (+4%). There are side benefits from the environmental viewpoint which could make carbonate leaching

even more attractive. This aspect is addressed in Chapter 6.0. Alternatives 1 and 2 for type B ore are not competitive and would not likely be justified.

The base case milling process for Ore C has comparable operating and amortized capital costs. Overall, milling costs are 47 per cent lower than type A milling costs. No alternative reviewed was economically practical.

5.0 HEALTH AND SAFETY

5.1 Introduction

Workers in uranium mills are exposed to a variety of hazards from machinery, chemicals, chemical and radioactive dusts and gases, as well as external radiation. The design of the mill and regular occupational health and safety monitoring programs are adjusted in consideration of these work-place hazards.

Radiation exposure and work-place hazards associated with the conventional sulphuric acid leaching process for all three ore types are well documented. Process alternatives that differ substantially from the conventional sulphuric acid leach may also differ significantly in terms of occupational health and safety. This section discusses the major health and safety differences between the various processes.

In general, hazards can be associated with the particular chemicals used, the by-products produced, and the complexity of the process. Industrial accidents for example are often measured in terms of man-hours worked, and it would be expected that total exposure to industrial accidents will vary directly with the establishment required to support the process.

5.2 Ore A - Process Alternatives

5.2.1 Alternative 1 - Pyrite Flotation

Pyrite flotation presents no significant additional health hazard in the milling operation provided adequate ventilation is supplied to remove mists from the flotation cells. Concentrate handling may or may not represent an increased hazard depending on the method of disposal. If placed directly underground there is a concern over the fire hazard, as pyrite is combustible. If mixed with backfill, there is little additional hazard other than the amount of slime in backfill would be slightly increased.

5.2.2 Alternative 2 - Acid Production

Roasting and acid manufacture are conventional processes which require control of dust and generated SO_2 gas. Because roasting is a dry process, a greater potential health hazard (dust) exists as compared with the base case sulphuric acid leach.

All of the generated sulphur dioxide is not converted to sulphuric acid, and a minor amount is released to the environment. For the acid plant specified, the release of SO_2 to the environment would amount to 1-2 t SO_2 per day.

5.2.3. Alternative 3 - Bulk Flotation

As with pyrite removal, bulk flotation does not represent a significant additional health hazard in the mill.

As described previously, the radioisotope flotation alternative does not operate at a high ratio of concentration and the activity of the concentrate is not significantly greater than that of the ore. With the provision of good ventilation above the flotation cells, radioisotope flotation does not appear to present a health hazard. However, should future research provide a method of substantially increasing the concentration of radium and thorium, a significant radiation hazard could exist, and flotation cells may require hooding and venting to a scrubber. Any concentration process will result in local hot spots of gamma activity which could lead to increased worker exposures.

Disposal of the concentrate does constitute a major concern. If underground disposal is selected, there are massive quantities of slimy material to be disposed. Given there is little excess space available in a mine, it is likely that those concentrates would be disposed in stopes near working areas. This can result in a safety hazard should a bulk head fail and release the slime. As well, potentially higher underground miner exposures can be expected due to the active placement and dewatering of these radioactive

sludges underground.

5.2.4 Alternative 4 - Thorium Recovery

The thorium recovery alternative is a minor modification to the basic acid leach circuit and does not represent a significantly different occupational hazard in the mill. Chemicals such as tributyl phosphate and alkyl phosphoric acid are irritants and toxic and good ventillation must be ensured.

The concentrate is handled as a dense slurry therefore minimal dusting should occur. The concentrate at 45 percent thorium and 26,000 Bq/g Th-230 needs to be handled with respect. Over the longer term as radium ingrowth occurs, the concentrate will require greater care and needs to be disposed with this in mind.

5.2.5. Alternative 5 - Autogenous Leach

Pressure sulphuric acid leaching, either for pyrite oxidation or for uranium oxidation does not increase health and safety problems. Minor hazards associated with pressure systems and oxygen gas are not materially greater than conventional leach hazards such as sodium chlorate handling.

5.2.6. Alternative 6 - EDTA Leach

Radium recovery from sulphuric acid leaching system with EDTA does introduce increased radium-related hazards into the mill. The EDTA circuit is physically very large and a large inventory of concentrated radium solution and sludge is present. The physical size alone results in increased opportunities for air-borne dust and elevated radon levels from process leaks and poor housekeeping.

The sludge from the filtration is a slurry, approximately 35 percent solids, with a radium concentrate of some 29,000 Bq/g of dry solids or 10,000 Bq/g of slurry. This compares with radium sludges in effluent treatment ponds in

Elliot Lake at 500 Bq/g of dry solids or 100 Bq/g of slurry. At 100 times these levels, gamma fields are substantial and shielding would be required. Production rates of 1200 m^3/a would mean approximately 20 standard barrels of this sludge would be produced daily. These concentrates would contain twice the radium content of pure pitchblende. Given the experience with the very high grade operations in Saskatchewan, worker health and safety aspects would require extensive measures.

Disposal underground could be readily accomplished as this volume is small. Other than the exposures associated with handling the sludge, the effects on the barrelled waste in the underground environment are anticipated to be minimal. However, the high specific activity of the material would dictate that extraordinary engineering and monitoring systems be in place.

5.2.7 Alternative 7 - HCl Leach

Hydrochloric acid solutions are much more hazardous to handle than sulphuric acid solutions because of the vapour pressure of HCl. At 70°C, a 20 g/L solution has an HCl vapour pressure of 0.01 mm Hg, or about 13 ppm. HCl vapour has a toxicity rating of 5 ppm compared to 1 ppm for chlorine and 10 ppm for hydrogen cyanide. It is evident tank venting and general ventilation standards become more important with an HCl leach, and there is increased risk of adverse health effects.

There are concerns over the handling of both the thorium and radium concentrates. The thorium concentrate handling aspects are similar to those addressed in section 5.4.4. Radium concentrate handling problems are greater than for EDTA leach as the concentrate is even more radioactive, however the slurry volume is reduced. At 180,000 Bq Ra-226/g, it is almost certain that special handling procedures and shielding will be required. This is true for both the mill operations and for concentrate containers. Without shielding at a 35% content and 60,000 Bq Ra-226/g of slurry, contact gamma levels in the order of 5 R/hr or greater can be anticipated for the barrelled waste. As such, barrelled waste will require substantial shielding and remote handling

systems that are typical of Type B, C low-level radioactive waste handling systems.

5.3. Ore B - Milling Alternatives

5.3.1 Alternative 1 - HCl-Leach

The concerns associated with HCl leaching of Ore B are analogous to Ore A except the concentrates are more radioactive. This is especially true for the thorium concentrate which would now increase in Th-230 content to 1,000,000 Bq/g.

5.3.2. Alternative 2 - HCl + Cl_2 Leach

Although the general comments applying to hydrochloric acid handling and managing radium concentrate for HCl leaching are appropriate, dry chlorination introduces other significant hazards. The chlorination roast creates a dust problem which must be minimized by maintaining a high standard of equipment maintenance and ventilation. Toxic volatile metal chlorides are formed in the roaster and converted to toxic metal oxides. Chlorine gas is highly toxic and requires careful handling. Knowledgeable staff must be on-hand in the event of a chlorine leak, and bulk chlorine storage is a potential hazard in the event of a fire or a spill.

For process reasons, oxygen must be effectively excluded from the dry chlorination roast, and for health and safety reasons, roaster gases must obviously be completely contained. Whether or not the roaster and gas handling equipment are maintained under a negative pressure, significant leakage into or out of the system can not be permitted and the design and operating/maintenance standards must be exemplary.

Relative to the base case process and other alternatives, chlorination represents so high a health and safety risk, that it is unlikely that the resultant environmental benefits could be justified.

5.3.3 Alternative 3 - Carbonate Leaching

Process solutions are generally less hazardous than acid solutions, but the uranium recovery circuit operates at a free caustic level of about 5 g/L, which requires safety precautions equivalent to those necessary for the handling of acid solutions. Within the mill, rubber and other lining materials are used only to a minor extent, solvent extraction is not applicable, and wood is not used as a construction material. Relative to an acid leach mill therefore, fire hazard during mill operation is minimal, and safety hazards after decommissioning due to structural deterioration would remain low for many years.

5.4. Ore C - Milling Alternatives

5.4.1 Alternative 1 - HCl Leach

The comments previously noted for Ores A and B are equally applicable.

5.4.2 Alternative 2 - HCl + Cl_2 Leach

The comments applying to HCl + Cl_2 Leach for Ore B are applicable. In addition this process will also recover arsenic trioxide, a very fine and highly toxic powder. Good ventilation and housekeeping priorities are necessary to control chemical exposure to arsenic. Arsenic trichloride, a highly poisonous substance, is likely to be formed sporadically.

Type C ore is found in remote areas of Saskatchewan. Any process that utilizes or produces highly toxic gases would require extraordinary engineering and management systems. Remote mines must rely on unskilled labour which at times would be incompatible with the requirements of chlorine technology. Without detailed analysis, it would appear that the costs and risks of such technology would far outweigh any short or long-term benefits.

6.0 ENVIRONMENTAL ASPECTS

6.1 Introduction

6.1.1 Overview

Alternative uranium processing technologies have several advantages and disadvantages. The environmental advantages can include the production of cleaner tailings that release lower levels of acidity, metals and radionuclides. The primary disadvantage is typically economics.

The basic physical facilities for environmental control include a tailings repository to store the solids and effluent treatment facilities to control waterborne contaminants. For the three basic ore types, the conceptual requirements can be quite different. The following is a general discussion on the typical requirements for disposal and treatment. These basic outlines will form the basis from which the alternative technologies can be compared.

6.1.2 Tailings and Effluent Management Requirements - Base Case

i) Type A Ore Practice

The base case facility will process 2.66 $x10^6$ tonnes of ore per year and will require a tailings basin to store a total of 56.7 x 10^6 tonnes of tailings generated during the 20 year lifetime of the facility. The tailings basin for this size facility would normally be located in a lake valley with a drainage basin of some 3-5 km^2 and a tailings surface area of 2 km^2. Tailings discharged at 45% solids could be expected to settle to an average density of 1.3 t/m^3. Assuming a 4 km^2 drainage basin and an annual specific runoff rate of 18 $L/s/km^2$ the total tailings effluent discharge would be 2.27 x 10^6 m^3/a from runoff and 1.92 x 10^6 m^3/a in excess water from the mill discharge for a total effluent discharge requiring treatment of 4.19 x 10^6 m^3/a. This assumes that for a valley disposal concept, the tailings remain saturated and as a result store a large quantity of the mill water discharged to the

tailings basin.

A modern treatment facility would include: pH adjustment, barium chloride addition in mechanically agitated contactors, followed by sand filters with continuous sludge disposal into the tailings basin.

ii) Type B Ore - Practice

The base case facility will process 0.594×10^6 tonnes of ore per year and will require a tailings basin to store a total of 12.9×10^6 tonnes of tailings generated over the 20 year lifetime of the facility. Assuming a Wollaston Lake, Saskatchewan location, the tailings area would most likely be a shallow valley or lake basin. Assuming an average depth of 10 m for the repository, and a specific gravity of 0.9 t/m^3 for in-place tailings, the tailings basin would cover approximately 1 km^2. The drainage basin could likely be limited to 2 km^2. Runoff from this region of Saskatchewan typically averages 10 $L/s/km^2$. Therefore the average runoff from the tailings facility would be 20 L/s or 0.63×10^6 m^3/a. Excess tailings water would add another 0.39×10^6 m^3/a for a total tailings effluent discharge requiring treatment of 1.02×10^6 m^3/a.

Effluent treatment would be likely be analogous to the type A ore facility.

iii) Type C Ore - Practice

The base case facility will process 0.115 million tonnes of ore per year and require a tailing basin to store a total of 2.7×10^6 tonnes of tailings generated over the 20 year lifetime of the facility. The tailings basin would most likely be above-ground engineered containment. Assuming a 10 m tailings depth and a dry density of the in-place tailings of 0.7 t/m^3, the tailings area would have a surface area of approximately 0.4 km^2 with no external drainage basin. Assuming tailings remain saturated and the natural runoff from the area averages 10 $L/s/km^2$, the total effluent requiring treatment would be 0.97×10^6 m^3/a, made up of $0.13 \times 10^6 m^3$ runoff and $0.84 \times 10^6 m^3$

excess tailings water.

Effluent treatment would include pH adjustment, iron addition for arsenic precipitation and either the standard barium chloride treatment for radium control or ion exchange. All sludges (and resin) would be disposed within the tailings basin.

6.1.3 Close-out Requirements - Base Case

i) General

Since 1983 in Canada, three uranium mines have been decommissioned and the properties reclaimed. These are Agnew Lake, Beaverlodge, and the Faraday Mine. The requirements for each were very different and future requirements are likely to be equally as variable. The basic regulatory requirements are negotiable and typically based upon safety and radiological hazards.

The major objective of the closeout plans will be to minimize the requirement for any long-term maintenance and institutional control. Based upon the fact that no acid-generating base metal mine has truly been able to prevent acidification without flooding of the tailings, the likelihood that the tailings areas for type A can be transformed into a "walkaway" situation is still open to question. Even type B and C tailings areas are not likely to be released from institutional and other long-term management controls.

ii) Type A Ore Tailings

The primary concern with base case tailings is acid production with radiological hazards of secondary importance. The four potential management concepts which could be considered are:

1) Grade and shape, and vegetate tailings surface. Provide for perpetual treatment for acidity and possibly radium removal.
2) Grade and shape, apply a shallow cover, vegetate and provide for

perpetual treatment for acidity and possibly radium removal. The primary difference to 1) is lower surface gamma levels and lower acid generation rates.

3) Apply a cover of non-acid-producing material to a sufficient depth to cause the water table to rise into the cover. This concept would minimize long-term acid production.

4) Apply a water cover.

Of the above options, the probable selections would be either 2) or 3). For this review it has been assumed that the low-grade tailings deposit for base case would be shallow-covered with 0.5 metre of soil or rock followed by grading, shaping and vegetation. Shallow cover would reduce surface gamma radiation levels to below 250 uR/hr and acid generation rates would be reduced. However, it is anticipated acid generation will occur for several hundred years making long-term treatment necessary.

iii) Type B Ore Tailings

Type B ore tailings are not acid generating but have higher surface gamma levels. Therefore a similar long-term management strategy to type A ore tailings is warranted. The medium-grade ore would also be graded, shaped, shallow-covered and vegetated. Interim treatment for radium removal would be expected with minimal long-term attention as acid generation will not occur.

iv) Type C Ore Tailings

Given the high activity level of the tailings, it is expected some form of special treatment of the tailings will be required. This could include the application of a multi-barrier cap or the relocation of the tailings to the pit. Long-term concerns relate more to intrusion than to environmental effects as little contaminant release to the environment is expected.

6.1.4 Environmental and Radiological Concerns

i) Type A Ore Tailings

The primary environmental concerns during operations are the effects on local water quality. Because of the large volume of discharge, typical effects include a general increase in total dissolved solids levels and potential toxic effects due to the presence of chemicals such as ammonia. Radiological concerns are related to radon releases from the tailings surface and downstream increases in radium levels in water supplies. As a general comment, none of these environmental or radiological aspects are typically of major concern during the operational phase.

Long-term environmental and radiological concerns are similar, other than for constituents such as ammonia which are no longer at elevated levels. The added concern relates to acid generation. Once the tailings become acidic, many heavy metals in the tailings become mobile as do the radionuclides of U-238 series. Of primary concern are thorium-230 and lead-210.

Given that treatment is provided, the effects of the facility are similar to the operating phase. The major problem is the high cost of treatment which could be required for several hundreds of years.

Without treatment, the major environmental effects would be acidification of the downstream watershed. Accompanying the acidification would be elevated levels of radionuclides and metals.

Long-term emissions of radon gas are a concern, but it is generally accepted radon is not a significant radiological pathway unless receptors are located very near to the tailings areas.

ii) Type B Ore Tailings

The environmental and radiological aspects of type B ore tailings are very

similar to type A ore tailings during the operating phase. Because treated effluent quality is similar, water quality effects per kilogram of uranium produced are lower as compared with the type A ore as far less water would be discharged. Radon emissions could be marginally higher than type A ore tailings because the tailings basins are typically shallower and contain higher levels of radium-226.

Long-term concerns (other than those related to intrusion) are greatly reduced. If covered, radon emissions would be reduced. The only significant contaminants expected in runoff are radium and possibly uranium. Assuming proper grading and drainage to minimize contact of runoff with the tailings, the downstream water quality can be expected to be at acceptable levels within a short distance of the tailings basin.

iii) Type C Ore Tailings

The environmental and radiological aspect of type C ore tailings are similar to type B ore tailings during the operating phase.

For the long term, if it is assumed that the multi-barrier engineered cap is applied, the only significant pathway for contaminants to leave the facility is via seepage to groundwater. The primary contaminants of concern are nickel, arsenic and radium-226. For anticipated low volumes of seepage that would occur it is most improbable that seepage will be a significant environmental or radiological problem, therefore no long-term treatment is expected.

6.2 Characteristics of Solid and Liquid Discharges

6.2.1. Solid Discharges

An overall summary of the characteristics for each of the alternative-process solid wastes is provided on Tables 6.2.1 and 6.2.2. The items listed include the annual tonnage, expected dry density, tailings volume, radionuclide

Table 6.2.1

ESTIMATED CHEMICAL AND RADIOLOGICAL CHARACTERISTICS - TYPE A,B, & C ORE AND TAILINGS

Ore Type	Alternative	Dry Density (t/m^3)	Annual Production (10^3 tonnes)	Annual Volume (10^3 m^3)	Solids Characteristics						
					U_3O_8 (%)	Thorium (%)	Ra-226 (Bq/g)	Th-230 (Bq/g)	FeS_2 (%)	As (%)	Ni (%)
A	ore feed	2.7	2664.7	-	0.09	0.025	9.3	9.3	5.6	-	-
A	Base Case Tailings	1.3	2836.4	2182	0.005	0.023	8.6	8.6(1)	5.3	-	-
A1	Pyrite Flotation Tailings	1.3	2686.6	2067	0.005	0.023	7.8	7.8	0.6	-	-
A2	Acid Production - Tailings	1.3	2830.4	2177	0.005	0.023	8.6	8.6	0.7	-	-
A3	Bulk Flotation Tailings	1.5	2036.5	1357	0.005	0.013	3.7	3.7	0.15	-	-
A4	Thorium Precip. Tailings	1.3	2751.6	2116	0.005	0.010	8.6	0.86 (2)	5.3	-	-
A5	Autogenous Leach	1.2	2998.1	2498	0.004	0.022	8.3	8.3	0.3	-	-
A6	EDTA Leach Tailings	1.3	2836.4	2182	0.005	0.023	2.6	8.6	5.3	-	-
A7	HCl leach Tailings	1.1	3018.7	2744	0.002	0.006	0.6	0.6	0.3	-	-
B	Ore Feed	2.7	595	-	0.4	0.002	41.5	41.5	0.19	0.02	-
B	Base Case Tailings	0.9	646.1	718	0.019	0.0018	38.2	38.2	0.17	0.02	-
B1	HCl Leach Tailings	0.8	721.8	902	0.004	0.0002	4.1	4.1	0.16	0.02	-
B2	HCl + CL_2 Tailings	1.3	595	458	0.02	0.002	4.1	41.5	<0.1	0.02	-
B3	Carbonate Leach Tailings	1.5	595	397	0.016	0.002	41	41	<0.1	0.02	-
C	Ore feed	2.8	115	---	2	0.02	207	207	1.9	1.5	2
C	Base Case Tailings	0.7	135.7	194	0.03	0.017	175	175	1.6	1.3	1.7
C1	HCl Leach Tailings	0.7	141.5	202	0.016	0.003	6.7	6.7	1.5	1.22	1.6
C2	HCl + Cl_2 Tailings	1.3	113.8	87.5	0.075	0.02	12.6	209	0.1	0.25	2

Notes

1) Assumed Th-230 reports in same size fraction as Ra-226. Therefore for physical separation process Th-230 activity approximately equal to Ra-226 activity.

2) Assumes Th-230 is dissolved in leaching 90% precipitated.

Table 6.2.2

SUMMARY OF CONCENTRATE VOLUMES & CHARACTERISTICS

Ore Type	Alternative	Concentrate Production (t/a)	Concentrate Volume (m^3/a)	Characteristics Ra (Bq/g)	Th (%)	Th-230 (Bq/g)
A1-	Pyrite Concentrate	150,000	100,000	16.7	0.03	16.7
A3-	Bulk Flotation	800,000	800,000	20.9	0.05	20.9
A4-	Thorium ppt	850	850	-	45	26,000
A6-	EDTA Ra concentrate	600	1,200	29,000	-	-
A7-	HCl Leach					
	-Ra concentrate	134	268	180,000	-	-
	-Th concentrate	1,070	1,070	-	45	22,000
	Total	1,204	1,338			
B1-	HCl Leach					
	-Ra concentrate	55	110	400,000	-	-
	-Th concentrate	22	22	-	45	1,000,000
	Total	77	132			
B2-	HCl + Cl_2 Leach					
	-Ra concentrate	55	110	400,000	-	-
C1-	HCl Leach					
	-Ra concentrate	11.5	23	1,800,000	-	-
	-Th concentrate	42	42	-	45	555,000
	Total	53.5	65			
C2-	HCl + Cl_2					
	-Ra concentrate	11.5	23	1,800,000	-	-
	-Arsenic(as As)	1440[1]				

Note
1 Arsenic concentrate is assumed to be saleable (contamination with radioactivity may make it unsaleable)

content (U_3O_8, thorium, Ra-226, Th-230), pyrite content and nickel and arsenic content.

One major assumption, made in the development of the tables is that Ra-226 and Th-230 are in secular equilibrium and will be dissolved coincidently with uranium from the same matrix in the ore. Therefore total thorium solubility is not directly proportional to Th-230 solubility as much of the thorium may be resistant to leaching in mineral species such as monazite.

The following general comments regarding Table 6.2.1 are provided:

Dry Density - Process alternatives can have a major effect on the in-place volume of tailings. Typically tailings density are a function of grain size, ore density and gypsum content. As a general rule dry density of the tailings decreases: as grain size decreases; ore density decreases; and gypsum content increases.

For type A ore tailings measured dry densities of 1.3 t/m^3 are typical. For bulk flotation, the slimes content of the tailings is reduced; therefore, the density is increased. For autogenous leaching, the gypsum content is marginally higher; therefore, the density is marginally reduced.

For type B ore tailings measured densities of 0.9 t/m^3 are typical. The variability between alternatives results primarily from changes to gypsum content which is essentially nil for carbonate leaching to 20 percent for HCl leaching.

Tailings Production - Tailings production quantities are equal to ore weight, less uranium product plus gypsum formed from acid addition and neutralization.

Tailings Volume - Volumes are simply calculated from the tailings production in tonnes divided by the estimated dry density.

Solid Characteristics - Based upon the ore characteristics and dilution resulting from chemical addition and product recovery.

The following general comments, regarding the concentrate/sludge characteristics summarized on Table 6.2.2, are provided:

Sludge/Concentrate Volumes

i) Type A Ore

- Pyrite flotation - 150,000 t/a at density of approximately 1.5 or 100,000 m^3/a.
- Bulk flotation concentrate - 800,000 t/a at a density of approximately 1.0 t/m^3 or 800,000 m^3/a.
- Thorium precipitate - 850 t/a approximately 1 t/m^3 or 850 m^3/a.
- EDTA $BaSO_4$ precipitate - 533 t/a of $BaCl_2$ is used. This would produce 1,200 m^3/a of radium slurry for disposal.
- HCl leach - i) Radium concentrate - requires about 120 t $BaCl_2$ at 0.1 g/l to precipitate radium. This produces 134 t of $BaSO_4$. At 35 percent solids this produces 268 m^3 of slurry for disposal. Thorium concentrate - Approximately 25 percent more thorium is recovered than in Alt. No.5; therefore, sludge volume is approximately 1,070 m^3/a.

ii) Type B Ore

- HCl Leach - radium concentrate would be similar to type A ore, as this is based upon uranium content. With higher pregnant solution concentration, the $BaCl_2$ demand is reduced and 55 t/a of sludge produced. Therefore, sludge production is 110 m^3/a. The thorium concentrate volume will be small as very little thorium is present. It was assumed a 45 percent concentrate could be produced. Total sludge production is 22 t/a or 22 m^3/a.

- $HCl + Cl_2$ Leach - Radium concentrate volume is again reduced as the volume of barren treated is reduced. The $BaCl_2$ demand is approximately 50 t/a producing 55 t/a of $BaSO_4$. At 35 percent solids the net slurry volume is 110.1 m^3/a.

iii) Type C Ore

- HCl Leach - radium concentrate volume is based on 10.2 t/a $BaCl_2$ use. This translates to 23 m^3/a of slurry. Thorium concentrate volume is based upon thorium concentrate at 45% solids. Total thorium removed approximately 19 t/a or 42 t/a concentrate. This equals approximately 42 m^3/a concentrate for disposal.
- $HCl + Cl_2$ Leach - radium concentrate is the same as for HCl leach, i.e. 23 m^3/a.

6.2.2 Liquid Discharges

The predicted chemical characteristics of liquid effluents from the tailings repository for the operational phase and long term are provided on Table 6.2.3.

The operational effluent levels provide for dilution of mill process water by precipitation and are the contaminent concentrations prior to treatment for radium and arsenic removal. Technology currently exists to reduce these contaminants to regulatory discharge levels.

The predicted long-term characteristics are for surface runoff from type A and B ore tailings repositories and for seepage from a type C ore tailings repository. The surface discharges are anticipated to be the major discharge pathways for type A and B ore tailings.

Type C ore base case facilities should only have a long-term groundwater release. Depending on the abandonment strategy selected, the alternative processes for type C ore may have a surface water discharge.

Table 6.2.3

TYPICAL CHEMICAL CHARACTERISTICS TAILINGS EFFLUENT DISCHARGE[3,4,5]

Ore Type Process	10^6 m^3/a[1]	Operational Ca	Operational SO_4	Phase Na	Phase Cl	U_3O_8	As	Ni	Ra-226[6]	Th-230[6]	10^6 m^3/a[2]	Long Term >200 years Ca	Long Term >200 years SO_4	U_3O_8	As	Ni	Ra-226[2]	Th-230[6]
A BaseCase	4.19	600	1600	205	315	<.1	<.01	<.1	30	1	2.27	100	240	<.1	<.01	<.1	1-10	<1
A1 - Pyrite Flotation	4.83	600	1600	170	260	<.1	<.01	<.1	24	1	2.15	100	240	<.1	<.01	<.1	1-10	<1
A2 - Acid Production	4.92	600	1600	170	260	<.1	<.01	<.1	26	1	2.27	100	240	<.1	<.01	<.1	1-10	<1
A3 - Bulk Flotation	4.55	600	1600	140	210	<.1	<.01	<.1	8.3	1	1.41	-	-	<.1	<.01	<.1	1-3	<1
A4 - Thorium Precipitation	4.07	600	1600	210	325	<.1	<.01	<.1	30	1	2.2	100	240	<.1	<.01	<.1	1-10	<1
A5 - Autogenous Leach	4.51	600	1600	-	-	<.1	<.01	<.1	30	1	2.6	100	240	<.1	<.01	<.1	<1	<1
A6 - EDTA Leach	18.5	600	1600	1480	380	<.1	<.01	<.1	5	1	2.27	100	240	<.1	<0.1	<.1	<1	<1
A7 - HCl Leach	2.12	600	1600	1140	6260	<.1	<.01	<.1	15	1	2.85	100-600	240-1600	<.1	<0.1	<.1	<1	<1
B Base Case	1.02	600	1600	500	665	1-2	<.01	<.1	40	1	0.63	600	1600	0.2	<0.01	<.1	20	<1
B1 - HCl Leach	0.51	600	1600	-	6500	<.1	<.01	<.1	25	1	0.79	600	1600	0.1	<0.01	<.1	<1	<1
B2 - HCl + Cl_2	0.29	600	1600	3040	39100	<.1	<.01	<.1	15	1	0.4	100	240	0.1	<0.01	<.1	<1	<1
B3 - Carbonate Leach	0.36	-	-	4760	-	5	<.01	<.1	4	1	0.35	-	-	0.2	<.01	<.1	<1	<1
C Base Case	0.97	600	1600	770	480	<.1	1-10	.05-2	50-150	1	0.32	600	1600	0.5	1-10	0.1-2	30	<1
C1 - HCl Leach	0.10	600	1600	-	640	<.1	1	.05	20-60	1	0.33	600	1600	0.5	1-10	0.1-2	<1	<1
C2 - HCl + Cl_2	0	600	1600	15000	45600[7]	<.1	1	.05	20-60	1	0.15	600	1600	0.5	<1	<1	0.1-2	<1

Notes

1 Based upon water balance and includes mill water discharge plus natural runoff less water stored as tailings porewater
2 Runoff flows based upon local rainfall/runoff data. Runoff for alternatives is ratioed with respect tailings volume as compared with base case
3 All levels prior to treatment
4 Where data were available, typical values are used. Otherwise judgement and best guesses were used to characterize effluents. This especially true for long-term predictions
5 All levels in mg/L unless stated
6 Levels in Bq/L
7 Porewater concentration as process water discharge is zero

In the long term (200 years and more), type A ores are assumed to have oxidized all their pyrite from the surface zone and therefore produce a near neutral discharge.

Runoff flows from each area are assumed to be those of the base case discussed in Section 6.2, times the ratio of the tailings volumes between the base case and process alternative. This assumption means that if the tailings volume is twice as large, the tailings basin is twice as large and therefore has two times the runoff. Effluent discharge flow is assumed to be total mill water discharge plus runoff less the amount of water stored in the tailings (assuming all tailings remain essentially saturated).

6.3 Estimated Costs

6.3.1 Tailings Disposal and Effluent Treatment Operating Costs

The annual operating costs for tailings and concentrate disposal, and effluent treatment are summarized on Table 6.3.1. These costs include:

- cost for tailings repository construction
- cost for sludge/concentrate disposal
- amortized capital cost for effluent treatment plant
- cost for effluent treatment chemicals
- overall maintenance and labour costs

As any disposal cost is site specific, generalized estimates based upon industry experience have been made. The details of the costing basis used are as follows:

- Tailings disposal costs have been based on repository construction. A cost of $\$0.50/m^3$ has been used for Type A ore. This is typical of Elliot Lake experience. Type B ore being disposed in shallower valleys would typically required more dyking, therefore a cost of $\$0.625/m^3$ (+25%) has been used. For type C ore tailings disposal in an above ground engineered/lined repository has been assumed. A cost of $\$5/m^3$ has been used for this type of disposal facility.

Table 6.3.1

OPERATING COSTS FOR TAILINGS DISPOSAL AND EFFLUENT TREATMENT

Ore Type	Alternative	Tailings Volume (10^6 m^3/a)	Tailings Disposal Cost (10^6 $)	Concentrate Volume (m^3/a)	Concentrate Disposal Cost (10^6 $/a)	Effluent Volume (10^6 m^3/a)	Effluent Treatment Plant (10^6/a)	Effluent Treatment Chemicals (10^6/a)	Operating and Maintenance Costs (10^6/a)	Operating Cost (10^6/a)
A	Base Case	2.182	1.09	0	0	4.19	0.53	0.06	0.33	2.01
A1	- Pyrite Flotation	2.067	1.04	100,000	1.0	4.82	0.58	0.07	0.31	3.00
A2	- Acid Production	2.177	1.09	0	0	4.92	0.59	0.07	0.33	2.08
A3	- Bulk Flotation	1.357	0.68	800,000	8.0	4.55	0.56	0.06	0.20	9.50
A4	- Thorium Precipitation	2.116	1.06	850	1.70	4.07	0.52	0.06	0.32	3.66
A5	- Autogenous Leach	2.498	1.25	0	0	4.51	0.56	0.07	0.37	2.25
A6	- EDTA Leach	2.182	1.09	1200	2.40	18.5	1.27	0.21	0.33	5.30
A7	- HCl Leach	2.744	1.37	1338	4.82	2.12	0.36	0.03	0.41	6.99
B	Base Case	0.718	0.45	0	0	1.02	0.23	0.05	0.14	0.87
B1	- HCl Leach	0.902	0.56	132	1.32	0.51	0.16	0.02	0.17	2.23
B2	- HCl + Cl_2 Leach	0.458	0.29	110	1.10	0.29	0.11	0.01	0.09	1.60
B3	- Carbonate Leach	0.397	0.25	0	0	0.36	0.13	0.02	0.08	0.48
C	Base Case	0.194	0.97	0	0	1.16	0.31	0.05	0.29	1.62
C1	- HCl Leach	0.202	1.01	65	0.65	0.29	0.14	0.01	0.30	2.11
C2	- HCl + Cl_2 Leach	0.088	0.44	23	0.23	0.14	0.09	0.01	0.13	0.9

BASIS:

Tailings Disposal Cost @$.50/$m^3$ type A ore, $.625/$m^3$ type B ore, $5/$m^3$ type C ore

Concentrate Disposal Costs - large volume wastes slurried to underground $10/$m^3$
- barrelled waste $2,000/$m^3$, concrete vaults
- small volume high activity wastes $10,000/$m^3$

Effluent Treatment Plant Cost - Based upon filtration plant cost of $6x$10^6$ for $8x10^6$ m^3/yr amortized @ 12% for 20 years. Cost increase or decrease by a factor of 0.5 as plant size doubles or halves i.e. cost/a = 0.23 (plant size 10^6 m^3/a)$^{.586}$. Add 25% for type C arsenic treatment.

Effluent Treatment Chemicals - 25 mg/L $BaCl_2$ - type A ore, 100 mg/L $BaCl_2$ types B and C ore, $BaCl_2$ cost $.50/kg, + 25% for type C ore arsenic removal

Operating and Maintenance Costs - $0.15/$m^3$ for type A, 0.19/m^3 for type B, $1.5/$m^3$ for type C ore tailings.

- All sludges or concentrates are disposed underground or in concrete vaults within the pit (where applicable). Underground disposal of pyrite concentrate or bulk flotation products would be by pumping into abandoned stopes. Costs of $\$10/m^3$ have been assumed. For radium and thorium concentrates, all these wastes would be barrelled and some would require shielding during handling. An allowance of $\$2000/m^3$ has been provided for underground disposal or disposal in concrete vaults within the pit for wastes with activity levels of <100,000 Bq/g. For small volume wastes with activity levels >100,000 Bq/g, which would require shielding, $\$10,000/m^3$ for disposal has been assumed.
- Effluent treatment plant costs will vary with size and complexity. A radium contact system followed by filters was assumed for all ore types. Arsenic removal is also provided for with type C ore. A simple regression equation was developed to determine the treatment plant cost versus plant size. For this relationship a cost of $6,000,000 was used for an eight million m^3/a capacity treatment plant. Then as plant size doubles or halves, treatment plant costs increase or decrease by a factor of 1.5. Therefore the amortized capital cost in million $/a =

  $$\text{cost } \$10^6/a = 0.23 \text{ (plant size million } m^3/a)^{.586}$$

  For arsenic removal 25 per cent was added to the capital costs for the treatment plant construction.

- Chemical costs were based upon $BaCl_2$ addition for radium removal and $FeCl_3$ addition for arsenic removal. $BaCl_2$ addition rates used were 25 mg/L for type A ore and 100 mg/L for type B and C. For type C ore 25 percent was added to the chemical costs for arsenic removal.

- Operating and maintenance costs are not well known. These costs include labour for relocating and repairing tailings lines, treatment plant operations, equipment rental, internal dykes, supplies, etc. An allowance of $\$0.15/m^3$ for type A, $\$0.19/m^3$ type B and $\$1.5/m^3$ for type C ore was provided.

. All capital costs are amortized at 12% over the 20 year operating life.

6.3.2 Abandonment Costs

The abandonment costs, which include a component for surface cover and vegetation, are summarized on Table 6.3.2.

For type A ore tailings, the basic requirement was assumed to be shallow cover (1/2 m) and vegetation. Shallow cover is applied to ensure a reduction in surface gamma levels to less than 250 uR/hr (AECB guideline for Beaverlodge decommissioning). Because process alternatives bulk flotation, EDTA leach and HCl leach will all reduce surface gamma through the removal of radium and/or thorium, no cover would be required.

Type B ore tailings were assumed to have similar requirements. Again HCl leach and the HCl + Cl_2 leach process alternatives eliminate the need for cover.

Type C ore tailings will require special treatment. For these tailings it was assumed that a multi-barrier engineered cap would be applied for the base case. The HCl leach reduces tailings radium activity from 175 Bq/g to 6.7 Bq/g therefore it is probable that direct vegetation is possible. The HCl + Cl_2 leach produces a lower volume of tailings and only removes the radium. Therefore shallow cover would likely be required.

As with the disposal costs, the abandonment costs have been amortized over 20 years at 12 percent.

6.3.3 Long-Term Treatment Costs

All processes are likely to require some short-term treatment but several process alternatives will not require treatment beyond the first few years after reclamation. The long-term treatment requirements will be for acidity control and radium removal, or in some cases for both. The treatment

Table 6.3.2

ABANDONMENT COSTS (W/O TREATMENT)

Ore Type	Alternative	Basic Requirements	Surface Area (ha)	Cover (10^6 m^3)	Cost ($10^6) Cover	Cost ($10^6) Vegetation	Total ($10^6)	Amortized Over 20 years ($10^6/a)
A	Base Case	shallow cover + vegetation	200	1.00	15	1.5	16.50	2.14
A1	- Pyrite Flotation	shallow cover + vegetation	189	0.95	14.18	1.42	15.60	2.02
A2	- Acid Production	shallow cover + vegetation	200	1.00	15	1.5	16.50	2.14
A3	- Bulk Flotation	vegetation	124	0	0	0.93	0.93	0.12
A4	- Thorium Precipitation	shallow cover + vegetation	194	0.97	14.55	1.46	16.01	2.08
A5	- Autogenous Leach	shallow cover + vegetation	229	1.15	17.18	1.72	18.90	2.45
A6	- EDTA Leach*	vegetation	200	0	0	1.50	1.50	0.19
A7	- HCl Leach	vegetation	252	0	0	1.89	1.89	0.25
B	Base Case	shallow cover + vegetation	100	0.50	7.5	0.75	8.25	1.07
B1	- HCl Leach	vegetation	126	0	0	0.95	0.95	0.12
B2	- HCl + Cl_2 Leach	vegetation	64	0	0	0.48	0.48	0.06
B3	- Carbonate Leach	shallow cover + vegetation	55	0.28	4.13	0.41	4.54	0.59
C	Base Case	engineered cap + vegetation	40	1.20	18	0.30	18.30	2.37
C1	- HCl Leach	vegetation	42	0	0	0.32	0.32	0.04
C2	- HCl + Cl_2 Leach	shallow cover + vegetation	18	0.09	1.35	0.14	1.49	0.19

BASIS:

Shallow cover - 0.5 m thick cover of soil or rock to reduce surface gamma to below 250 uR/hr ($15/$m^3$)

Vegetation - grade, shape, and seed @ $7500/ha

Engineered cap - 3 m thick multi-barrier engineered cap @ $15/$m^3$.

requirements and costs are summarized on Table 6.3.3.

Treatment for acidity removal will be required only for type A ore tailings. Costs for acidity removal are based upon $25,000/a for labour, $25,000/a for power and supplies and $225,000/a for lime addition (based upon 1000 mg/L acidity).

Radium removal costs are flow dependent. A labour allowance of $25,000/a and a chemical cost of $15,000 for every million m^3 treated with 25 mg/L of $BaCl_2$ was provided. .

The annual long-term treatment cost was then converted into an operational cost in order to determine the effect on the cost of uranium production. This cost was first turned into a capital investment cost (i.e. how much money must be invested at 10% to provide a perpetual fund equal to the long term disposal cost which increases annually at a 5% rate of inflation.) Once this capital investment was determined it was then amortized over 20 years at 12% to determine the prorated cost on an annual operating basis.

6.3.4 Total Costs for Tailings Management

The total costs for tailings management are summarized on Table 6.3.4. These costs are simply the sum of the operations costs, abandonment costs and long-term treatment costs.

The base case cost for type A ore is $5.01 x $10^6/_a$ or $2.22/kg U_3O_8 produced. Alternative processes range in cost from $1.93/kg to $4.31/kg U_3O_8.

The base case cost for type B ore is $2.03 x $10^6/_a$ or $0.90/kg U_3O_8 produced. Alternative processes range from $0.47/kg to $1.02/kg U_3O_8.

The base case cost for type C ore is $3.99 x $10^6/_a$ or $1.77/kg U_3O_8. Alternative processes 2 and 3 were 0.95/kg and $0.50/kg respectively.

Table 6.3.3

LONG-TERM TREATMENT COSTS (10-200 YEARS AFTER CLOSURE)

Ore Type	Alternative	Flow (10^6 m^3/a)	Treatment Req'd: Acidity	Treatment Req'd: Radium/Arsenic	Annual Cost ($10^6)	Investment (1) ($10^6 $)	Amortized @ 12% over 20 years ($10^6 $)
A-0	Base Case	2.27	Yes	Yes	0.334	6.64	0.86
A-1	Pyrite Flotation	2.15	No	Yes	0.058	1.16	0.15
A-2	Acid Production	2.27	No	Yes	0.059	1.18	0.15
A-3	Bulk Flotation	1.41	No	Yes	0.046	0.92	0.12
A-4	Thorium Precipitation	2.20	Yes	Yes	0.333	6.66	0.86
A-5	Autogeous	2.60	No	Yes	0.064	1.28	0.17
A-6	EDTA Leach	2.27	Yes	No	0.275	5.50	0.71
A-7	HCl Leach	2.86	No	No	0.000	0.00	0.00
B-0	Base Case	0.63	No	Yes	0.034	0.68	0.09
B-1	HCl Leach	0.79	No	No	0.000	0.00	0.00
B-2	HCl + Cl_2 Leach	0.40	No	No	0.000	0.00	0.00
B-3	Carbonate Leach	0.35	No	No	0.000	0.00	0.00
C-0	Base Case	0.13	No	No	0.000	0.00	0.00
C-1	HCl Leach	0.13	No	No	0.000	0.00	0.00
C-2	HCl + Cl_2	0.06	No	No	0.000	0.00	0.00

Acidity Removal - allow $25,000 labour + $225,000 for lime + $25,000 utilities = $275,000/a for base case
Radium Removal - allow $25,000 labour + $15,000/$10^6$ m^3 treated

1 Investment required at 10% return to maintain a perpetual fund with an annual expenditure growing at the rate of inflation (5%)

Table 6.3.4

TOTAL COSTS
TAILINGS DISPOSAL AND EFFLUENT MANAGEMENT

Ore Type	Alternative	Operating Costs ($\$10^6$/a)	Abandonment Cost ($\$10^6$/a)	Long-Term Treatment ($\$10^6$/a)	Total Cost ($\$10^6$/a)	$/kg U_3O_8	% Difference
A	Base Case	2.01	2.14	0.86	5.01	2.22	-
A1-	Pyrite Flotation	3.00	2.02	0.15	5.17	2.29	+3
A2-	Acid Production	2.08	2.14	0.15	4.37	1.93	-13
A3-	Bulk Flotation	9.50	0.12	0.12	9.74	4.31	+94
A4-	Thorium Precipitaton	3.66	2.08	0.86	6.60	2.92	+32
A5-	Autogenous	2.25	2.45	0.17	4.87	2.15	-3
A6-	EDTA Leach	5.30	0.19	0.71	6.20	2.74	+24
A7-	HCl Leach	6.79	0.25	0.00	7.24	3.11	+40
B	Base Case	0.87	1.07	0.09	2.03	0.9	-
B1-	HCl Leach	2.23	0.12	0.00	2.35	1.02	+13
B2-	HCl + Cl_2 Leach	1.60	0.06	0.00	1.66	0.73	-18
B3-	Carbonate Leach	0.48	0.59	0.00	1.07	0.47	-47
C	Base Case	1.62	2.37	0.00	3.99	1.77	-
C1-	HCl Leach	2.11	0.04	0.00	2.15	0.95	-46
C2-	HCl + Cl_2 Leach	0.9	0.19	0.00	1.09	0.50	-72

6.4 Comparison of Alternatives

6.4.1 General

A summary table of the major features of the process alternatives is provided on Table 6.4.1. These include the effects of process alternatives on tailing area requirements, quantities of concentrate, specific activity, acid generation potential, radon emissions (both short and very long term), future uses, operational and long-term contaminant loadings, long-term treatment requirement and cost. The highlights of Table 6.4.1 are discussed for each ore type below.

6.4.2 Type A Ore Milling Alternatives

Alternative 1 - Pyrite Flotation

Pyrite removal is an economically competitive alternative from a environmental viewpoint. Costs for concentrate disposal are offset by the reduced costs for long-term treatment. In addition, there are minor long-term benefits the radiological composition of the tailings. Pyrite concentrates tend to also concentrate radium, therefore the activity of the tailings and the coincident radon releases are marginally reduced. One minor disadvantage is the increased water use which results in marginally higher operational loadings of dissolved salts and radium.

For pyrite removal to be an acceptable practice, concentrate must be either used to produce acid (as in alternative 2) or it must be disposed in an acceptable fashion. If underground disposal cannot be accommodated, an acceptable alternative would need to be found.

Alternative 2 - Acid Production

Acid production will reduce the costs for tailings disposal and as such is an attractive alternative. The primary long-term benefit is the elimination of

COMPARISON OF ALTERNATIVE PROCESSES TO THE BASE CASE FROM AN ENVIRONMENTAL PERSPECTIVE

Consideration	Base Case	1 Pyrite Flotation	2 Acid Production	3 Bulk Flotation	4 Thorium Precipitation	5 Autogenous Leach	6 EDTA Leach	7 HCl Leach
	- TYPE A ORE -							
Tailings Volume/Area	-	-5%	-	-38%	-1%	+14%	-	+26%
Other Solid Wastes								
. type	-	pyrite	-	float conc.	thorium	-	radium	thorium & radium
. quantity	-	150,000 t/a	-	800,000 t/a	850 t/a	-	600t/a	1204 t/a
. saleability	-	no	-	no	unlikely	-	no	unlikely
. disposal cost	-	high	-	very high	very high	-	high	very high
Specific Activity of Tailings								
. Ra-226 (Bq/g)	8.6	7.8	8.6	3.7	8.6	8.3	2.6	0.6
. Th-230 (Bq/g)	8.6	7.8	8.6	3.7	0.86	8.3	8.6	0.6
Acid Generation Potential	yes	no	no	no	yes	no	yes	no
Radon Emissions								
. short term	-	-14%	-	-69%	-1%	+10%	-70%	-91%
. very long term	-	-14%	-	-69%	-90%	+10%	0	-91%
Future Uses								
. housing	no	no	no	no	no	no	no	possible
. recreational	possible	possible	possible	possible	possible	possible	possible	yes
. construction material	no	no	no	no	no	no	no	possible
Liquid Discharges Volume								
. operational	-	+15%	+17%	+9%	-3%	+8%	+342%	-49%
. long term	-	-5%	-	-38%	-3%	+15%	-	+26%
Contaminant "Loadings"								
- Operational								
. $CaSO_4$ + NaCl	-	+11%	+14%	+2%	-2%	-13%	+559%	+162%
. Ra-226	-	+15%	+17%	+9%	-3%	+8%	+342%	-49%
. Th-230	-	+15%	+17%	+9%	-3%	+8%	+342%	-49%
. Ni + As	-	-	-	-	-	-	-	-
- Long Term								
. Ra-226	-	-5%	-	-38%	-3%	+15%	>-90%	>-90%
. Th-230	-	-5%	-	-38%	>-3%	+15%	-	U-90%
. Ni + As	-	-	-	-	-	-	-	-
Treatment Requirement at Abandonment for								
. Acidity	yes	no	no	no	yes	no	yes	no
. Radium	likely	likely	likely	likely	likely	likely	likely	no
Overall Costs % for Tailings	-	+3	-13	+94	+32	-3	+24	-40

Table 6.4.1 continued

Consideration	- TYPE B ORE - Base Case	1 HCl Leach	2 $HCl + Cl_2$ Leach	3 Carbonate Leach	- TYPE C ORE - Base Case	1 HCl Leach	2 $HCl + Cl_2$ Leach
Tailings Volume/Area	-	+26%	-36%	-45%	-	+4%	-55%
Other Solid Wastes							
. type	-	thorium + radium con.	radium conc.	-	-	thorium + radium conc.	radium conc.
. quantity	-	77 t/a	55 t/a	-	-	53.5 t/a	11.5 t/a
. saleable	-	unlikely	unlikely	-	-	unlikely	unlikely
. disposal cost	-	high	high	-	-	high	high
Specific Activity of Tailings							
. Ra-226 (Bq/g)	38.2	4.1	4.1	41	175	6.7	12.6
. Th-230 (Bq/g)	38.2	4.1	41.5	41	175	6.7	209
Acid Generation Potential	no	no	no	no	low	no	no
Radon Emissions							
. short term	-	-86%	-90%	-40%	-	-96%	-97%
. very long term	-	-86%	0	-40%	-	-96%	0
Future Uses							
. housing	no	no	no	no	no	no	no
. recreational	possible	yes	no	no	no	no	no
. construction material	no	no	no	no	no	no	no
Liquid Discharges Volume							
. operational	-	-50%	-72%	-65%	-	-75%	eliminated
. long term	-	+25%	-37%	-44%	-	+3%	-53%
Contaminant "Loadings"							
- Operational							
. $CaSO_4$ + NaCl	-	+29%	+1192%	-27%	-	-64%	eliminated
. Ra-226	-	-50%	-72%	-65%	-	-75%	eliminated
. Th-230	-	-50%	-72%	-65%	-	-75%	eliminated
. Ni + As	-	-	-	-	-	-75%	eliminated
- Long Term							
. Ra-226	-	-95%	-95%	-96%	-	-97%	-97%
. Th-230	-	-90%	-37%	-44%	-	97%	-53%
. Ni + As	-	-	-	-	-	+3%	-90%
Treatment Requirement at Abandonment for							
. Acidity	no	no	no	no	no	no	no
. Radium	likely	no	no	no	yes	no	no
Overall Costs %	-	+13	-18	-47	-	-46	-72

the need to treat for acidity control. Other tailings characteristics are essentially unchanged. The primary disadvantages would include higher water use and increased SO_2 emissions.

Alternative 3 - Bulk Flotation

The bulk flotation alternative is very expensive but does have some positive benefits. The high cost is primarily the result of the vast quantity of concentrate which requires disposal. In addition to the high cost, it is highly likely that underground disposal of the concentrate would not be possible. This is not to say this alternative would not be applicable during the last one or two years of operation.

Bulk flotation results in no long-term acid generation, reduced surface gamma levels, reduced rodon emissions and reduced levels of radium in the discharge water over the long-term.

Alternative 4 - Thorium Precipitation

Thorium precipitation is an expensive process and is not likely justifiable unless the thorium concentrate is saleable. The major increase in cost from management is for concentrate disposal. The benefits are only accrued over the very long term as radium-226 will still be present in the tailings until it decays away.

Alternative 5 - Autogenous Leaching

Autogenous leaching from waste management viewpoint has reduced costs and is a very attractive process alternative. One drawback is the increased tailings volume which was assumed to be produced because the excess acid produced from the pressure oxidation would require neutralization. Therefore the tailings have more gypsum and are likely to settle to a lower bulk density thus requiring a larger tailings basin.

Alternative 6 - EDTA Leach

The EDTA leach process is more expensive for environmental management. This is primarily because of the high cost for concentrate disposal. The advantages include reduced radium activities which in the long term would be re-established by the ingrowth of radium from the parent Th-230.

There are many operational drawbacks which include high water use, high salt loadings to the environment, and potential occupational concerns related to concentrate management. If this process could be simply implemented it could merit some consideration as short term add-on during the late years of operation. However this is not the case, and as such this alternative does not appear to be economically competitive.

Alternative 7 - HCl Leach

Hydrochloric acid leaching produces an environmentally cleaner tailings which overall are economically attractive from the waste management perspective. The increased costs for concentrate disposal are more than offset by the reduced costs for long-term tailings management. The tailings production volume is greater but the low activity of this material makes it virtually innocuous. The tailings would have minimal restriction on future use and may be acceptable for use in construction material.

The major concern would be the management of the thorium and radium concentrate and the health and safety aspects during operations.

6.4.3 Type B Ore Milling Alternatives

Alternative 1 - HCl Leaching

The HCl leach process is marginally more expensive for waste management but the tailing characteristics are greatly improved. The only major concern relates to the occupational health and safety aspects of handling the highly

radioactive radium and thorium concentrates.

Alternative 2 - HCl + Cl_2 Leaching

The chlorination process produces a reduced tailings volume at a reduced overall waste management cost. The primary disadvantages are higher operational salt loadings and disposal of the highly radioactive concentrate. Over the very long term, Ra-226 levels will re-establish therefore one would question whether this process as described is worthwhile.

Alternative 3 - Carbonate Leach

Carbonate leaching has substantially reduced waste management costs. This is primarily because this process produces a dense tailing material which requires less disposal volume. In addition with a smaller tailings surface area radon emissions are reduced. As well carbonate leaching uses less water and is expected to have lower operational and long term radium discharges.

6.4.4 Type C Ore Milling Alternatives

The advantages of the HCl leach and HCl + Cl_2 leach are the same as discussed for Ore B. The major additional advantage is that waste management costs are even more reduced for this ore type.

7.0 SUMMARY

7.1 Overall Costs

The total cost for milling and waste management are summarized in Table 7.1.1. Although waste management costs are often reduced for process alternatives, without exception alternative processes result in overall higher costs.

For type A ore the base case cost is \$25.44/kg U_3O_8. This cost does not include mining, head office costs, profit, and is based upon amortizing capital over the entire 20 year operating life which would not be done in practice. These additional costs could easily double the base case production costs leaving no flexibility to handle additional costs when one considers that world uranium prices are currently in the range of \$50/kg U_3O_8. Certainly the process alternatives 6 and 7 (EDTA and HCl Leach) are uneconomic and could not be practically considered.

For type B ore the base case cost is \$17.87/kg U_3O_8 or some 30 per cent less than type A ore. Similarly one would also expect lower mining costs. Carbonate leaching of this ore type is cost competitive and as such could be considered as a viable economic alternative. HCl leaching and chlorinates at 1.5 and 2 times the total costs could not be reasonably justified.

For type C ore the base case cost is \$14.16/kg U_3O_8 or 44 per cent less than for Type A ore and 20 percent less than Type B ore. As with Type B ore, the economics for alternative processes HCl leaching or chlorination are prohibitive.

A breakdown of the cost component for the milling and waste management (operation and long term) are provided on Table 7.1.2. The following general comments are noted.

- waste management costs represent from 5.1-12.5 per cent of the total processing costs

Table 7.1.1

OVERALL COST SUMMARY

Ore Type	Alternative	Production (10^6 kg U_3O_8/a)	Capital Cost ($/kg U_3O_8)	Operating Cost ($/kg U_3O_8)	Disposal Cost ($/kg U_3O_8)	Total Cost ($/kg U_3O_8)	Per cent Differen
A	BASE CASE	2.26	7.93	15.29	2.22	25.44	-
A1-	Pyrite Flotation	2.26	8.46	16.72	2.29	27.48	+8
A2-	Acid Production	2.26	10.82	15.93	1.93	28.68	+13
A3-	Bulk Flotation	2.26	8.75	19.12	4.31	32.18	+26
A4-	Thorium Precipitation	2.26	8.79	16.97	2.92	28.68	+13
A5-	Autogenous Leach	2.26	10.38	16.10	2.15	28.63	+13
A6-	EDTA Leach	2.26	12.31	40.26	2.74	55.31	+117
A7-	HCl Leach	2.33	14.87	30.48	3.11	48.46	+90
B	BASE CASE	2.26	6.39	10.58	0.90	17.80	-
B1-	HCl Leach	2.31	9.27	16.62	1.02	26.91	+51
B2-	HCl + Cl_2 Leach	2.26	13.23	22.41	0.73	36.37	+104
B3-	Carbonate Leach	2.26	6.50	11.01	0.47	17.98	+1
C	BASE CASE	2.26	5.73	6.66	1.77	14.16	-
C1-	HCl Leach	2.27	6.54	12.75	0.95	20.24	+43
C2-	HCl + Cl_2 Leach	2.20	7.87	15.70	0.50	24.07	+70

Table 7.1.2

FACTORS IN OVERALL COST

Processing Cost - Base Cost

	Milling Cost (%)		Disposal Cost (%)			
	Capital	Operating	Operations	Long-Term	Total (%)	Total ($/kg U_3O_8)
A - Base Case	31.2	60.1	3.5	5.2	100	25.44
B - Base Case	35.8	59.2	2.1	2.9	100	17.87
C - Base Case	40.5	47	5.1	7.4	100	14.16

- long-term waste management costs are greater than the operational tailings disposal cost. This is a very important factor
- long-term waste management costs are \$1.33/kg U_3O_8 for type A ore, \$0.51/kg U_3O_8 for type B ore and \$1.05/kg U_3O_8 for type C ore.

7.2 Type A Ore Alternatives

Of the process alternatives considered, only alternative 7, HCl leach produced a clean tailings. This process is uneconomic and in fact processing costs, excluding mining, profit, head office etc., are approximately equal to current world prices for uranium. EDTA leaching is even more cost prohibitive with fewer environmental benefits.

Based upon this review, it is very doubtful that any of the alternatives would be adopted for a greenfield site. However, this does not preclude the implementation of an add-on process during the latter years of operations to reduce long-term costs. Two practical examples are presented below, based on the following:

> Assume pyrite removal or bulk flotation is adopted during the last two years of operation so that the surface layer of tailings is not acid generating. Does the increased benefit offset the cost?

The additional capital cost for pyrite removal is \$9,300,000. The incremental operating cost for 2 years is \$6,460,000, for a total additional milling cost of \$15,690,000. With pyrite removal long-term treatment costs would be reduced and allowance would be necessary for concentrate disposal. Concentrate disposal costs are \$2,000,000 for two years. The long-term treatment investment is \$1,160,000. Capital costs for abandonment would be the same as for the base case or \$16,500,000. Therefore the total cost for pyrite removal from the surface cover including abandonment would be \$35,350,000. This compares with the base case cost of \$16,500,000 for capital costs for reclamation plus an investment of \$6,630,000 for long-term treatment, for a total of \$23,130,000. Based on the above noted costs, if

pyrite removal was to be carried out only during the last two years versus during the entire 20 year lifetime of the facility, the estimated cost would be \$0.27/kg U_3O_8 as compared with \$2.03/kg U_3O_8. These estimates are based on the entire 20 year production of U_3O_8.

Similarly the additional capital cost for a bulk flotation circuit is \$14,400,000. The incremental cost to operate the circuit for two years would be \$17,320,00. Concentrate disposal would be \$16,000,000, capital cost for abandonment \$1,460,000 and long-term treatment cost \$1,160,000 (similar to pyrite removal cost), for a total cost of \$50,340,000. This is \$0.60/kg U_3O_8 more than the base case cost. In comparison if bulk flotation was used from the start of the operation it would cost \$6.74/kg U_3O_8 more than the base case cost.

For both options significant additional cost savings would be possible if the capital equipment could be salvaged and if underground disposal costs could be reduced. For example if the capital equipment could be salvaged at 20% of its installed value and if underground disposal costs were reduced to \$2/m^3 the total costs for pyrite removal would be marginally reduced to \$0.24 kg U_3O_8 and bulk flotation cost significantly reduced to \$.38 kg U_3O_8.

Both of these revised options are still major costs to any operator. But, these options better meet the objective for decommissioning which is to minimize the need for human intervention. Certainly if tailings areas are major acid generators, long-term treatment and intervention is necessary to maintain acceptable water quality conditions down stream. With pyrite removal or bulk flotation long-term treatment for acidity should not be required, therefore the only contaminant of concern is likely radium-226. Given that runoff levels from existing areas are typically in the 1-10 Bq/L range, little downstream dilution is necessary to meet the drinking water criterion level of 1 Bq/L.

7.3 Type B Milling Alternatives

Of the alternatives reviewed, only the HCl leach milling alternative approaches the production of "clean tailings". The HCl + Cl_2 leach produces a low radium tailings that still contains thorium-230. Given the cost for chlorination is so much greater, this alternative warrants no further consideration.

Even with the hydrochloric acid leaching circuit, tailings will still contain elevated levels of radium and thorium. Although these levels have been reduced from about 40 Bq/g to 4 Bq/g these materials could in no way be left for unrestricted use. These materials still are radioactive; could not be used for building materials; and the areas would not be suitable for housing.

There are also some negative aspects with respect to concentrate handling. Great concern is expressed over handling and processing high-grade ores yet these ores contain orders of magnitude less radioactivity than do the radium concentrates from this process (approximately 400,000 Bq/g).

Given the cost difference of \$9.04/kg U_3O_8 to implement HCl leaching (i.e. more than \$400,000,000 over the life of the facility) as compared with the minor benefits, this process alternative is not justifiable.

Carbonate leaching is economically viable and has some potential environmental benefits. These primarily arise from the fact that tailings densities are greater and as such the net area necessary for tailings disposal is smaller.

7.4 Type C Milling Alternatives

As with type B ore, no milling alternative can produce "clean tailings". As for the same reasons discussed for Ore B, there would appear to be no justification for the adoption of these alternatives. The alternatives do reduce the detrimental levels but given the very high cost it is unlikely that any case could be developed to warrant their implementation.

8.0 REFERENCES

1. Raicevic, D., 1979. "Decontamination of Elliot Lake Uranium Tailings." CIM Bulletin, August.

2. Pakkala, T.E., 1978. "Pilot Study of Pyrite and Radioisotope Removal by Flotation." Denison Mines Limited, CANMET OSQ 78-00143, November.

3. Sirois, L.L., Cristorici, M.A., Raicevic, D. "Preconcentration of Uranium Ore."

4. James F. MacLaren Limited, 1978. "Environmental Assessment of One Proposed Elliot Lake Uranium Mines Expansion, Volume 4." April.

5. Vermeulen, L.W., 1965. "Recovery of Thorium from Uranium Process Solutions." Presented at the 94th Annual Meeting of AIME, February 1965.

6. Dasher, John, 1972. "A Conceptual Design for the Leaching of Sulphidic Uranium Ores Without Reagents." Minerals Science Engineering, October.

7. Harrison, V.F. and Smith, H.W., 1960. "Air Oxidation Pressures Leaching of Sulphide-Bearing uranium Ores." Dept. of Mines and Technical Services, Ottawa. Internal Report EMI 60-2, May.

8. Vezina, J.A. and Gow, W.A., 1969. "Some Design Aspects of the Pressure-Oxidation Acid Leaching of a Canadian Uranium Ore." Department of Energy, Mines and Resources Technical Bulletin TB110, December.

9. Pendreigh, R., 1978. "Successful Trials or Pressure Leaching of Uranium." Coal Gold and Base Minerals of Southern Africa, September.

10. Bovey, J.H. and Stewart, L.N., 1978. "Pressure Leaching of Uranium Bearing Witwatersrand Ores." The South African Institute of Mining and Metallurgy.

11. Yagnik, S.K., Baird, M.H.I., and Banerjee, S., 1981. "An Investigation of Radium Extraction from Uranium Mill Tailings." Hydrometallurgy, 7 pp. 61-75.

12. Ryan, R.K., Levins, D.M., 1980. "Extraction of Radium from Uranium Tailings." CIM Bulletin, October, pp. 126-133.

13. Nixen, A., et al., 1983. "Radium Removal from Elliot Lake Uranium Mill Solids by EDTA Leaching." Hydrometallurgy, Vol. 10, pp. 173-186.

14. Nixon, A., et al., 1983. "The Recovery of Radium and EDTA from Radium Leach Solutions." Hydrometallurgy, Vol. 11. pp. 33-45.

15. Seeley, F.G., 1976/77. "Problems in the Separation of Radium from Uranium Ore Tailings." Hydrometallurgy, Vol. 2, pp. 249-263.

16. Skeaff, S.M., "Proposed Route for Nitric Acid Leaching of Uranium Ore and Removal of Ra-226 of EDTA." CANMET Internal Report No. MSL-INT-78-101.

17. Hague, K.E., 1979. "Hydrochloric Acid Leaching of an Elliot Lake Uranium Ore." CANMET Report ERP/MSL 79-59 (OP), June.

18. Hague, K.E., 1981. "Hydrochloric Acid Leaching of Rabbit Lake Uranium Ore." CANMET Report ERP/MSL 81-138 (TR), November.

19. Hague, K.E., 1982. "Hydrochloric Acid Leach of an Elliot Lake Uranium Concentrate." CANMET Report ERP/MSL 82-72 (TR), July.

20. Borrowman, SR., Brooks, P.T., 1975. "Radium Removal from Uranium Ores and Mill Tailings." USBM R/8099.

21. Ritcey, G.M., Weng, E.W., "Solvent Extraction of Uranium from a Chloride Leach Liquor." CANMET Report ERP/MSL 84-57 (J)

22. C.E. Lummus, 1980. "Evaluation of Chlorine/Chloride Based Processors for Uranium Ores." Minerals Division, CANMET, Contract Report No. OSQ80-00030, November.

23. Campbell, M.C., et al., 1981. "Chloride Metallurgy for Uranium Recovery, Conceptual Costs." CANMET Report ER/MSL 81-125 (OP) October.

24. Skeaff, J.M., "Chlorination of Uranium Ore for Extraction of Uranium, Thorium and Radium and for Pyrite Removal." CIM Bulletin, V01. 27, No. 801, pp. 120-125.

25. Skeaff, J.M., Laliberte, J.J., 1980. "Continuous Hihg Temperature Chlorination of Uranium Ore." CANMET Report ERP/MSL 80-48 (OP), March.

26. A.H. Ross Files

27. Ryan, A.D., Hurst, F.J., and Seeley, F.G., 1977. "Nitric Acid Leaching of Radium and Other Significant Radionuclides from Uranium Ores and Tailings." Oak Ridge National Laboratory, ORNL/TM-5944, August.

28. Skeaff, J.M., 1975. "Nitric Acid Leaching of Uranium Ore." CANMET Report No. MRP/MSL 75-165 (TR) August.

29. Skeaff, J.M., 1978. "Proposed Route for Nitric Acid Leaching of Uranium Ore and Removal of Ra-226 by EDTA." CANMET Internal Report No. MSL-INT 78-101, November.

30. Lummus, 1981. "Economic Evaluation of Preconcentration of Uranium Ores: Final Report." CANMET Report OSQ80-00120, April.